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ANALYSIS OF ENERGY SYSTEMS AND PERFORMANCE IMPROVEMENT OF A KRAFT PULPING MILL

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RÉSUMÉ

Cette étude a pour objectif d'augmenter l'efficacité énergétique d'une usine existante de fabrication de pâte à papier Kraft. Le principal moyen mis en œuvre pour atteindre ce but consiste à développer de nouvelles conceptions optimisées du procédé de fabrication, en d'autre termes il est question ici d'augmenter la récupération interne de chaleur et le taux des fermetures des circuits hydrauliques afin de réduire la consommation énergétique de l'ensemble du procédé. Dans un premier temps il est nécessaire de développer un modèle numérique de l'usine afin d'obtenir les bilans de masses et d'énergie du procédé a l'aide d'un logiciel de simulation de procédés chimiques appelé CADSIM plus® . Ensuite la simulation a été validée par comparaison avec le fonctionnement réel de l'usine sur les paramètres importants dont la consommation d'eau et de vapeur par des mesures in situ avec le personnel de l'usine. Les écarts relatifs de production et de consommation d'eau et de vapeur sur la totalité des flux de l'usine n'excèdent pas 5 %.

La simulation a été caractérisée et comparée aux valeurs moyennes de consommation des usines canadiennes du secteur des pâtes et papiers. Par ailleurs ; les réseaux de vapeur et d'eau de l'usine ont été établis clairement ainsi que les bilans de masse associés. Les profils de température et de consistance de la pâte à papier le long de la ligne de production ont été tracés afin d'identifier les inefficacités énergétiques liées aux points de mélange non isothermiques. L'étude comprend l'analyse des contraintes techniques de l'usine basée sur une approche systématique et documentée. Un manuel technique d'analyse des contraintes a été rédigé, il peut être appliqué à n'importe quelle usine de production de pâte à papier. Les effets potentiels en termes d'économie d'énergie liés aux différents niveaux de contraintes ont été étudiés à l'aided'une analyse globale comprenant l'aspect technique et économique. En terme de re-conception totale du procédé, la ligne A n'a présenté une réduction des consommations que de l'ordre de 2 %, aucune différence remarquable n'a pu être relevée sur la ligne B. Théoriquement la diminution de consommations obtenue sur la ligne A est de 22 % en re-conception totale et 20 % en re-conception partielle. En considérant différentes conceptions du réseau d'échangeur, il est possible d'atteindreune diminution de la consommation de 17 % en re-conception totale et 15 % en re-conception partielle. Pour la ligne B, d'après les courbes composites la diminution théorique de consommation maximale est de 24 % et 16 % d'après la conception du réseau d'échangeur existant.

Une analyse économique a été réalisée à partir du réseau d'échangeurs de la ligne A, elle montre que dans le cas d'une re-conception partielle du procédé, on peut atteindre un temps de retour brut sur investissement de 2,1 ans et de 3,1 ans dans le cas d'une re-conception totale. Ces résultats sont justes si la production de vapeur et la consommation de combustible sont réduites. On peut donc dire qu'il est économiquement rentable d'imaginer une re-conception partielle ou totale de la ligne A de l'usine. Pour la ligne B, la re-conception partielle du réseau d'échangeurs conduit à un temps de retour brut de 3,6 ans si on assure une diminution de la production de vapeur. La solution qui consiste à augmenter la production de vapeur de l'usine afin d'en accroitre la production d'électricité s'est relevée économiquement non rentable pour les lignes A et B.

ABSTRACT

An energy study was done with the objective of improving the energy efficiency of an existing Kraft pulp mill. The improvements have been achieved by developing optimized process designs for the energy systems. The first step was to develop Mass and energy models of the mill on CADSIM plus® software. Second, the model was validated by examining water and steam results and other major parameters. The discrepancy in total steam and water production and consumption was less than 5%. The configuration of the model has been validated directly with the mill staff.

The mill has been characterized and benchmarked against Canadian industry average. In addition, steam and water networks have been built and mass balances around these two systems were done. The temperature and consistency profiles of pulp and water tanks were plotted and inefficiencies due to non isothermal mixing in the process have been identified.

Constraint analysis was performed on the overall mill based on a systematic and documented approach. A set of guidelines have been developed in order to customize the constraint analysis process to any pulp and paper mill. The effect of different constraint levels such as grassroot and retrofit on energy savings has been studied by examining the total savings and economic data. In terms of grassroot and retrofit approaches, it was apparent that in line A the grassroot approach savings were more by 2% while for line B the difference was insignificant. Theoretical savings based on the composite curves for line A were 22% in grassroot and 20% in retrofit. Based on the different heat exchanger network designs, it was possible to achieve 17% savings in grassroot and a maximum of 15% in retrofit. For line B, theoretical savings based on the composite curves were 24% and the potential savings based on the heat exchanger network design was 16%.

An economic analysis was carried where by the heat exchanger networks of line A, show that for the retrofit case, a simple payback period of 2.1 years is achievable while for the grassroot case a simple payback period of 3 years is achievable. This is the case when steam production and fuel consumption are reduced. Therefore, one can say that it is economically viable to design either in grassroot or retrofit constraint level for line A. For line B, the retrofit heat exchanger network was built with a simple payback period of 3.6 years if reducing the steam production is the chosen scenario. Increasing the steam production to produce more electricity was not an economically feasible scenario for both line A and line B.

TABLE OF CONTENTS

ACKNOWLEDGMENTS III
RÉSUMÉ IV
ABSTRACT VI
TABLE OF CONTENTS VII
LIST OF TABLES XI
LIST OF FIGURES XIII
NOTATION XVII
LIST OF APPENDIXES XVIII
CHAPTRE 1 INTRODUCTION AND CONTEXT 1
1.1 Problem statement 1
1.2 Context 1
1.3 Objectives 2
1.3.1 General Objective 2
1.3.2 Specific Objectives 2
1.4 Hypothesis 2
1.4.1 Original scientific hypotheses of contribution (OSHC) 2
1.4.2 Originality Justification: 3
1.4.3 Refutability: 3
1.5 Structure and organization 4
CHAPTRE 2 LITERATURE REVIEW 5
2.1 Kraft Process 5
2.2 Methods and techniques 8
CHAPTRE 3 METHODOLOGY 11

3.1 Project Phases....11
3.2 Breakdown of the phases....12
3.3 Definition of the methods, techniques and tools....13
CHAPTRE 4 SIMULATION MODEL, VALIDATION AND CHARACTERIZATION.....15
4.1 Introduction....15
4.2 Simulation Model....16
4.3 Validation of the model....18
4.3.1 Water Validation - Line A....19
4.3.2 Water Validation - Line B....20
4.3.3 Steam Validation - Line A....21
4.3.4 Steam Validation - Line B....22
4.3.5 Steam validation - Line A + B....23
4.3.6 Other Parameters validation - Line A:....24
4.3.7 Other Parameters validation - Line B....25
4.4 Characterization of the model....26
4.4.1 Steam network....27
4.4.2 Water network....31
4.4.3 Benchmarking....35
4.4.4 Key Performance Indicators....38
4.4.5 Pulp line profiles....39
4.4.6 Water tanks profiles....43
CHAPTRE 5 GUIDELINES TO CONSTRAINT ANALYSIS IN A KRAFT MILL............45
5.1 Introduction....45
5.2 Constraint analysis guidelines....46

5.2.1 In-depth knowledge of the process....47

5.2.2 Identifying and screening the heat transfer points 48

5.2.3 Categorizing the type of heat transfer point....48

5.2.4 Listing possible energy savings projects....59

5.2.5 Identifying possibilities for grassroot and retrofit representations....60

5.2.6 Evaluating possible scenarios prior to building composite curves 63

5.2.7 Building the composite curves 66

5.3 Results....67

5.3.1 Water system– Integrated or separated 67

5.3.2 Line A and B – Integrated or Separated....69

5.3.3 Grassroot approach vs. Retrofit approach....72

5.3.4 Identifying all potential energy saving projects....79

5.3.5 Building the refined composite curves....81

5.3.6 Line A refined composite curves 83

5.3.7 Line B refined composite curves....84

5.3.8 Summary of refined composite curves....85

CHAPTRE 6 HEAT EXCHANGER NETWORKS AND ENERGY SAVING PROJECTS 86

6.1 Introduction:....86

6.2 Building the existing heat exchanger network....87

6.2.1 Existing heat exchanger network – Line A....87

6.2.2 Existing heat exchanger network – Line B 90

6.3 Evaluation of energy violations in the existing network....92

6.3.1 Energy violations - line A 92

6.3.2 Energy violations - line B....94

6.4 Building the new heat exchanger networks....................96

6.4.1 Line A - Retrofit – Low savings....................98

6.4.2 Line A - Retrofit – Medium savings....................104

6.4.3 Line A – Grassroot - High savings....................113

6.4.4 Summary of heat exchanger networks –Line A....................125

6.4.5 Line B – Retrofit – Medium savings....................126

6.4.6 Summary of heat exchanger network - Line B Retrofit medium....................137

6.5 Economic Analysis:....................138

6.5.1 Heat exchanger networks - Line A....................140

6.5.2 Heat exchanger network - Line B....................141

CHAPTRE 7 CONCLUSION AND RECOMMENDATIONS....................142

7.1 Conclusions....................142

7.2 Recommendations....................143

BIBLIOGRAPHIE....................144

ANNEXES....................147

LIST OF TABLES

Table 4-1: Major Model information 17

Table 4-2: Water validation - Line A 19

Table 4-3: Water validation - Line B 20

Table 4-4: Steam Production - Line A 21

Table 4-5: Steam consumption - Line A 21

Table 4-6: Steam production - Line B......22

Table 4-7: Steam consumption - Line B 22

Table 4-8: Steam production and consumption - Line A and B......23

Table 4-9: Other parameters validation - Line A 24

Table 4-10: Other parameters validation - Line B 25

Table 4-11: Steam balance - Line A......28

Table 4-12: Steam balance - Line B......30

Table 4-13: Breakdown of water balance – Line A 32

Table 4-14: Overall water balance - Line A......32

Table 4-15: Breakdown of water consumption - Line B......34

Table 4-16: Overall water balance - Line B 34

Table 4-17: Key Performance Indicators 38

Table 5-1: Initial energy saving projects - line A......59

Table 5-2: Initial energy saving projects - line B......59

Table 5-3: summary of the composite curves results......79

Table 5-4: Line A potential energy saving projects......80

Table 5-5: Line B potential energy saving projects 80

Table 5-6: Summary of refined composite curves results - Line A 85

Table 5-7: Summary of refined composite curves results - Line B 85

Table 6-1: List of Existing heat exchanger - Line A..89

Table 6-2: List of Existing heat exchangers - Line B 91

Table 6-3: Cross pinch violations - Line A 93

Table 6-4: Cross pinch violations - Line B 94

Table 6-5: Summary of energy saving projects - Retrofit low savings.......................................103

Table 6-6: Potential energy savings - Retrofit medium savings 112

Table 6-7: Summary of potential energy savings – Grassroot A 124

Table 6-8: Summary of savings at different constraint levels – Line A 125

Table 6-9: Summary of potential energy saving projects – Retrofit medium B 136

Table 6-10: Summary of heat exchanger network - Retrofit medium B.....................................137

LIST OF FIGURES

Figure 2-1: Kraft process overview....5

Figure 3-1: Breakdown of the methodology phases....12

Figure 4-1: Overall view of the model....17

Figure 4-2: Steam network - Line A....27

Figure 4-3: Steam network - Line B....29

Figure 4-4: Water network - Line A....31

Figure 4-5: Water network - Line B....33

Figure 4-6: Benchmarking - Thermal consumption....35

Figure 4-7: Benchmarking - Water consumption....36

Figure 4-8: Benchmarking - Effluent production....37

Figure 4-9: Temperature profile - Line A....39

Figure 4-10: Consistency profile - Line A....40

Figure 4-11: Temperature profile - Line B....41

Figure 4-12: Consistency profile - Line B....42

Figure 4-13: Warm water and Hot water tanks profile - Line A....43

Figure 4-14: Warm water and hot water tanks profile - Line B....44

Figure 5-1: Overview of constraint analysis and heat exchanger network methodology....47

Figure 5-2: Categorization of constraint type....50

Figure 5-3: Constrained Indirect steam Injection - 1a....51

Figure 5-4: Non constrained indirect steam injection -1b....52

Figure 5-5: Constrained direct steam injection – 1c....53

Figure 5-6: Non constrained direct steam injection 1d....54

Figure 5-7: Indirect Heat transfer between different process streams - 2....55

Figure 5-8: Energy in effluents and gases - 3 55

Figure 5-9: Non isothermal mixing in pulp line - 4 57

Figure 5-10: Non isothermal mixing in water tanks - 4 58

Figure 5-11: Retrofit approach 61

Figure 5-12: Grassroot approach 62

Figure 5-13: Existing water network for line A and line B 65

Figure 5-14: Thermal power for Water system - Integrated or separated 67

Figure 5-15: Total area for Water system - Integrated or separated 68

Figure 5-16: Capital cost for Water system - Integrated or separated 69

Figure 5-17: Thermal power for Line A and B – Integrated or Separated 70

Figure 5-18: Total area for Line A and B – Integrated or Separated 70

Figure 5-19: Capital cost for Line A and B – Integrated or Separated 71

Figure 5-20: Method of obtaining grassroot and retrofit schematic 72

Figure 5-21: Line B - Grassroot vs. Retrofit composite curves 73

Figure 5-22: Grand composite curve - Line B Process vs. water 74

Figure 5-23: Grand composite curve - Close up Line B 75

Figure 5-24: Line A - Grassroot vs. Retrofit composite curves 76

Figure 5-25: Grand Composite Curve - Process vs. water Line A 77

Figure 5-26: Grand composite curve - Close up Line A 78

Figure 5-27: Line A - Grassroot After and before refinement 83

Figure 5-28: Line A - Retrofit After and before refinement 83

Figure 5-29: Line B - Grassroot After and before refinement 84

Figure 5-30: Line B - Retrofit After and before refinement 84

Figure 6-1: Existing heat exchanger network - Line A 88

Figure 6-2: Existing heat exchanger network - Line B 90

Figure 6-3: Criss cross violations chart - Line A 93

Figure 6-4: Criss cross violations chart - Line B 95

Figure 6-5: Project 1 - Bleach Heater...98

Figure 6-6: Project 2- Brown Heater..99

Figure 6-7: Project 3 - Boiler Air Heater 100

Figure 6-8: Project 4- Deaerator Mae up water...101

Figure 6-9: Project 5 - Injection 1: Washer 15...102

Figure 6-10: Project 1 - Bleach Heater...104

Figure 6-11: Project 2 - Brownstock heater 105

Figure 6-12: Project 3 - Boiler Air Heater 106

Figure 6-13: Project 4 - Make up Water 107

Figure 6-14: Project 5 - Injection 1 Bleaching Washer 15..108

Figure 6-15: Project 6 - Injection 2 – Bleaching Washer 35..109

Figure 6-16: Project 7 – Injection 3 – Washer 45 110

Figure 6-17: Project 8 – Injection 4 – Washer 55 111

Figure 6-18: Project 1& 2 – WW to Washing and Bleaching......................................113

Figure 6-19: Project 3 – Hot water to bleaching 114

Figure 6-20: Project 4 – Hot water to Machine...115

Figure 6-21: Project 5 & 6 – Hot water to Recausticizing 1 & 2 116

Figure 6-22: Project 7 – Hot water to Bleaching 117

Figure 6-23: Project 9 – Make up water (Deaerator) 118

Figure 6-24: Project 10 – Injection 1 – Washer 15 119

Figure 6-25: Project 11 – Injection 2 – Washer 35 120

Figure 6-26: Project 12 – Injection 3 – Washer 45 121

Figure 6-27: Project 13 – Injection 4 – Washer 55 122

Figure 6-28: Project 14 – Boiler Air Heater....................123

Figure 6-29: Project 1 – Boiler Air Heater B 126

Figure 6-30: Project 2 – Make up water (Deaerator) B....................127

Figure 6-31: Project 3 – Non isothermal mixing in dilution conveyer 128

Figure 6-32: Project 4 – Non isothermal mixing in white water tank....................129

Figure 6-33: Project 5 – Bleach heater and Direct condenser....................130

Figure 6-34: Pre Project – Elimination of violations in cold blow cooler and green liquor cooler 131

Figure 6-35: Project 6 – Injection 1 – Washer 15 B....................132

Figure 6-36: Project 7 – Injection2 – Washer 35 B....................133

Figure 6-37: Project 8 – Injection 3 – Washer 45 B....................134

Figure 6-38: Project 9 – Injection 4 – Washer 55 B....................135

Figure 6-39: Steam savings economical scenarios Line A 140

Figure 6-40: Steam savings economical scenarios....................141

NOTATION

BL	Black Liquor
BLH	Black Liquor Heater
BSH	Brown stock heater
CBL	Cold blow liquor
CBC	Cold blow cooler
CL	Cooking Liquor
DVSE	Dust Vent scrubber exchanger
FS	Flashed steam
FSC	Flashed steam condenser
FW	Fresh Water
GL	Green liquor
GLC	Green liquor cooler
HW	Hot Water
ICC	Indirect Contact Cooler
SC	Surface Condenser
SWH	Shower water heater
WBL	Weak black liquor
WL	White liquor
WW	Warm Water

LIST OF APPENDIXES

Appendix 1 - Simulation model, validation and characterization …………………………147

1.1. Steam network tables

1.2. Injected steam

1.3. Water network tables

Appendix 2 - Guidelines for constraint analysis in a Kraft mill ………………………..….159

1.1. The complete list of all direct injection steam constraints

1.2. water system data for retrofit and grassroot

1.3. Lists of streams used to build the initial composite curves

1.4. Actual heating requirement

1.5. Information and equations used to evaluate the total area and capital cost

1.6. Water system – separated or integrated composite curves

1.7. Both lines - integrated vs. separate composite curves

1.8. List of streams for non isothermal mixing projects

1.9. List of effluents after refinement

Appendix 3 - Heat exchanger networks and energy saving projects …………………….175

1.1. List of streams used for building the heat exchanger network

1.2. Economic analysis data

1.3. List of modified heat exchanger networks and projects for line A

1.4. List of modified heat exchanger networks and projects for line B

CHAPTRE 1 INTRODUCTION AND CONTEXT

1.1 Problem statement

Low paper prices and demand, external competition and high energy costs have caused economic problems for the Canadian pulp and paper industry [1]. As a result, significant efforts are being undertaken to transform the pulp and paper industry into an efficient and profit oriented industry. A pioneer solution that addresses this issue is the retrofitting of biorefineries into existing mills. The implementation of biorefinery options could increase the profitability of the mills by creating a sustainable process with high value secondary products. A key step to be undertaken before the implementation of a biorefinery option is the optimization of a mill with respect to energy and water consumption. Increasing the efficiency of the mill would be achieved in a methodological way that involves a detailed analysis of the energy systems. This could result in steam savings projects scenarios. The promising projects are going to be compared, and analyzed based on technical economic constraints in order to select the potential projects. In addition, the excess steam could be integrated into the biorefinery to insure maximum reutilization of these utilities.

1.2 Context

This project is part of the BioKrEn project whereby 3 mills are being optimized in terms of water and energy. A different biorefinery option will be proposed for each of the optimized mills. The focus will be on the energy optimization of a western Canadian mill. The methodology in this project is adopted from the unified methodology presented in Mateos 2009[2]. In the unified methodology, the method for constraint analysis is based on experience and not a systematic approach for analyzing the constraints and extracting the data to perform pinch analysis. There is a need for a concrete set of guidelines to be followed to achieve realistic theoretical energy targets. A significant section of this project will be involved in developing guidelines for an efficient way to build composite curves and achieve energy targets. These guidelines will build a platform for future energy analysis projects.

1.3 Objectives

1.3.1 General Objective

To improve the energy efficiency of an existing Kraft pulp mill by developing optimized process designs for the energy system.

1.3.2 Specific Objectives

1. Develop and validate a base case simulation for a Kraft mill.
2. Characterize and evaluate the performance of the mill and locate inefficiencies in the process.
3. Perform constraint analysis to develop guidelines for screening between different constraint levels and find theoretical energy targets.
4. Propose potential energy saving projects and develop heat exchanger networks based on techno-economic factors.

1.4 Hypothesis

1.4.1 Original scientific hypotheses of contribution (OSHC)

OSHC: By following the proposed methodology, finding process Inefficiencies and analyzing them will lead to scenarios for reducing steam consumption.

OSHC 1: By using data from a mill, it is possible to develop and validate the simulation.

OSHC 2: By benchmarking the mill, it is possible to identify the inefficiencies in the process.

OSCH 3: By extracting the data in grassroot representation, it is possible to increase the theoretical energy saving scope and maintain a similar capital cost to retrofit approach.

OSCH 4: By developing heat exchanger networks, it is possible to obtain and evaluate the scope of potential savings and type of projects.

1.4.2 Originality Justification:

OJ: This global and local energy study has not been done on this mill before

OJ 1: Development and validation of the simulation has not been done on CADSIM plus ® software.

OJ 2: This base case mill has not been characterized and evaluated in a detailed manner.

OJ 3: The role of data extraction in grassroot and retrofit on the energy savings has not been analyzed in the literature.

OJ 4: The best project for energy savings for the mill has not been determined yet.

1.4.3 Refutability:

RF: If by following the methodology proposed, inefficiencies were not found or the savings were too small to be economically feasible, the hypothesis will be refuted.

RF 1: To be refuted if simulation data differs from the mill data (+/- 10 %)

RF 2: The hypothesis will be refuted if benchmarking doesn't lead to finding inefficiencies.

RF 3: The hypothesis will be refuted if the grassroot data extraction has no effect on the scope and cost of energy savings when compared to retrofit data extraction.

RF 4: The hypothesis will be refuted if the savings are too small or the projects proposed are not applicable due to economic constraints.

1.5 Structure and organization

In chapter 2, the literature review to support the methods and the findings of this project will be presented

In chapter 3, the overall methodology for energy optimization will be presented. The main focus will be on explaining the different steps in the methodology.

In chapter 4, the results and the techniques used in developing the simulation model, validation, and characterization will be presented.

In chapter 5, constraint analysis results and ideas will be discussed thoroughly and the outcomes of that section will be presented.

In chapter 6, the potential energy saving projects and the heat exchanger networks will be presented. In addition, the economic analysis of the projects and heat exchanger networks will be included.

In chapter 7, the conclusions and recommendations will be presented.

CHAPTRE 2 LITERATURE REVIEW

2.1 Kraft Process

Chemical pulping was first achieved in the early 1860's by treating wood with caustic soda. In 1867, advances in the chemical pulping lead to the use of calcium bisulphate as a pulping agent. In the late 1800's, sulphite pulping had become the dominant pulping method. Towards the end of the century, Kraft pulping was developed. The Kraft process is a chemical process that uses wood chips as feed material to produce pulp and paper products. Kraft pulping produces strong fibres even though it has shorter cooking times than other chemical pulping processes. The advantage of this type is that it is compatible with most types of wood. Wood chips are primarily composed of cellulose, hemicellulose, and lignin. Lignin is a highly organic compound that acts as the glue that keeps the other components intact. Harsh alkaline conditions are required to break down the lignin and free the hemicelluloses and cellulose for the pulp or paper making process. The chemical charge involved is a highly alkaline mixture (pH ~14) of sodium hydroxide (NaOH), sodium sulphide (Na_2S) and other sodium compounds used to degrade lignin [3].

The Kraft process has dominated the pulp and paper industry since 1940's because Kraft mills are able to recycle almost all of the pulping chemicals and have well integrated heat recovery systems. Kraft remains till this day the leading chemical pulping method. The Kraft process overview is presented in figure 2-1. Eight major departments in a Kraft process are needed to produce the final product. These departments are:

Figure 2-1: Kraft process overview

Pretreatment

In this department, wood chips are screened to remove large wood pieces and any impurities such as rocks or sand from the feed stream. In addition, the wood chips need to be heated with steam to 130 °C in order to replace the trapped air in chips with steam condensate[3]. This will result in a more efficient impregnation stage due to easier diffusion between impregnation liquor and water. The heated pulp is then sent to the digester in the cooking section.

Cooking

Cooking the pulp is essential to break down and dissolve the lignin to release the cellulosic and hemicellulosic material to produce pulp and paper. Another term for this process is called delignification. This process requires strong alkaline conditions, high temperature and pressure. The alkaline solution used is called white liquor and it mainly consists of NaOH and Na_2S. The cooking temperature is around 160 °C and the pressure is around 1100 kPa[3]. After the cooking stage in the digester, hemicelluloses and cellulose form the pulp while lignin is dissolved in the liquor to form black liquor. The mixtures is washed and cooled down to around 80 °C before it is sent to the washing department.

Washing

Pulp is washed using water or filtrate from bleaching in a cascading manner to separate the lignin from the pulp. The separated lignin forms the black liquor whereby it is sent to the evaporators department. The washed pulp continues to the bleaching department.

Bleaching

In the bleaching department, relatively clean pulp is bleached using chemicals to brighten the pulp into a white color. The bleaching chemicals used are mainly chlorine dioxide dissolved in water, peroxide, and sodium hydroxide. There are many consecutive stages of chemical injection and washing. Hot water, warm water, and fresh water are used to wash the pulp. Acidic effluents and alkaline effluents are released during the bleaching. Finally, clean and white pulp is sent to the forming department.

Forming

The bleached pulp is pressed and formed into sheets before the dryer. The water released from pressing the low consistency pulp is used as a source of water in bleaching. The pressed sheets are usually dried at two stages; low pressure steam is directly injected in the first stage to increase the temperature of the pulp while in stage two, medium pressure steam is used in an indirect contact dryer. The pulp is produced at a consistency of 90 %.

Evaporators

The black liquor from the washing department enters the evaporators at an average dissolved solid concentration of 19 %. The black liquor is concentrated using low pressure steam in an indirect evaporators operating under vacuum to reach a dissolved solid concentration of 40-50%. The concentrated black liquor is sent to the boilers to be burnt for energy production.

Steam Plant

The black liquor with high concentration of lignin is sent to the recovery boiler to be burnt. The energy from the black liquor is used to produce high pressure steam. The produced steam is sent to a turbine to produce electricity and different levels of steam to be used in the process as a heat source. The spent delignification chemicals at the bottom of the furnace are called the smelt. Smelt is highly concentrated in sodium carbonate and calcium carbonate. The smelt is sent to the chemical recovery department.

Chemical Recovery

Smelt from the recovery boiler is dissolved in water to produce green liquor. Green liquor undergoes chemical reactions to produce the white liquor. The reactions involve the transformation of calcium carbonates into calcium oxides under high temperatures. The calcium oxides are then reacted with sodium carbonates to produce white liquor. The produced white liquor is an alkaline that is mainly composed of sodium hydroxide and sodium sulphide. White liquor is sent to the digesters department thus closing the chemical recovery loop.

2.2 Methods and techniques

The techniques and tools used for process integration and optimization are many. Some include automated optimization paths while other techniques follow a manual path to increase the efficiency of a process [4]. A preliminary step that must be undertaken before any energy study is the development of a reliable base case model that represents a long term average of the operating conditions of a process [5]. The model should include general yet detailed data on the mass and energy balances of the process. Configuration and output data such as steam and water consumption need to be validated to ensure the reliability of the energy analysis [6].

Analyzing the energy and water systems of the mill could be a tedious process if it is not approached in a structured manner. Therefore, evaluating the mill from a general and local perspective will help in channelling efforts to address points of concern in an efficient manner. In other words, the process of benchmarking will reduce the time it takes to locate and analyze points of inefficiencies in an energy audit [7]. Representing the steam and water networks is an essential preliminary step in benchmarking [2]. This will help in locating and identifying the production, utilization and post utilization systems of water and steam. Deepening the understanding of these systems will lead to an easier identification of inefficiencies. Another tool that was mentioned in the literature is the comparison of energy, water, and electricity consumption of a mill with the Canadian average and best practice mills [8, 9]. An energy audit of a Scandinavian mill was performed whereby the mill was compared to the average Scandinavian industry and a state of the art mill. The results showed that excess steam and electricity consumption was higher in the dryers and the evaporators departments of the specific mill. This lead to a preliminary appraisal stating that inefficiencies occurred in both departments and further analysis is required to pinpoint the exact location of the inefficiencies [10].

In the literature, other indicators have been used to evaluate the performance of the process. Calculating and comparing key performance indicators of a mill with the Canadian average was reported to be helpful in identifying general inefficiencies [6]. Some of the key performance indicators include boilers efficiencies, flue gases energy losses, and condensate return [11].

Temperature screening tools are presented in the literature to analyze the consistency and temperature profile across the pulp line as well as the mixing temperatures in water tanks [12]. The outcome of this tool results in the identification of non-isothermal mixing points and direct heat transfer points in the pulp line and water tanks. By using a temperature vs. enthalpy variation criterion, the relevant points are screened and included in the energy analysis section.

Data handling and extraction is a crucial step that plays a big role on the energy analysis results. The extracted data could have multiple representations and each representation will have an effect on the practicality of the energy savings results. Two main distinct approaches that have not been applied to a Kraft process yet are the grassroot and retrofit representation of data. In the retrofit approach, the streams required are extracted based on the actual conditions and constraints in the mill. On the other hand, grassroot extraction disregards the constraints in the process whereby data is extracted based on final targets. Equipment such as heat exchangers and tanks are not considered as a constraint in the grassroot approach and thus the flexibility of matching hot and cold streams and building an optimized heat exchanger networks will increase [12].

The thermal pinch analysis is a technique that was developed in the late 1980's to improve the exchange of energy in a process [13]. The principle is fairly simple whereby the hot streams (to be cooled) and cold streams (to be heated) are extracted and plotted on a composite curve. The outcome of this curve includes the minimum heating and cooling requirement, and the possible internal heat recovery based on the design of a new heat exchanger network. The typical savings in the pulp and paper industry based on this principle are between 15%-30% [14]. The water pinch analysis is based on the same principles as thermal pinch analysis. The cold streams become the water sinks and the hot streams become the water sources [15]. The two sets of streams are plotted on a composite curve and the outcome indicates the maximum theoretical water reutilization in a given process. The use of this technique was developed in a manual mode but with the technology advancement, automated optimized algorithms were developed [16]. It is reported that the use of this technique can result in 20-40 % of fresh water reduction [14].

Thermal pinch analysis has been used in the literature to develop energy savings projects. In a case study based on a mill located in eastern Quebec, energy savings were around 20% [2]. On the other hand in the oil and gas industry, energy savings were up to 30 % [14].

A structured approach to retrofitting of heat exchanger networks have been proposed in the literature[17]. A set of general guidelines were developed to aid the engineer in achieving realistic energy targets. The information was very general and therefore a set of guidelines are needed to the pulp and paper industry. A fairly simple procedure for the design of heat exchanger networks was presented in the literature[18]. Reiteration of the pinch rules and how they should be applied to a new design is covered in this article. Moreover, the idea of assessing the pinch violations was described as well. A more focus article that discusses the retrofitting of an air heating system for a paper making industry represented the practicality of using the simple pinch methods[19]. It was proven that energy savings could be achieved by modifying the heat recovery system. Three heat exchanger networks were designed and evaluated based on their economic prospect.

In order to evaluate different retrofit scenarios in the existing heat exchanger networks, a graphical method is used to examine the possible energy saving scenarios[20]. The focus was on the location of heaters and coolers whereby the closer the coolers and heaters to the pinch, the more cost effective design will be. In the case study, practical results showed the by eliminating Criss-cross violations, higher energy savings are achieved[21]. More work has been done on the elimination of Criss-cross violations whereby a hot stream at high temperature levels is used to heat a cold stream at low temperatures without crossing the pinch[22]. The elimination of these violations as well as the use of an automated program, resulted in heat exchanger network design which have higher energy savings and requires much less time to be obtained.

CHAPTRE 3 METHODOLOGY

3.1 Project Phases

The project consists of four phases. Many techniques, tools and methods are used in each phase to meet the objectives. The phases are listed below:

Phase 1- Development of the simulation and validation:

Phase one will involve the development of a validated base case simulation. Validation of the configuration as well as the output/input data will be carried with the mill staff.

Phase 2 – Characterization:

The second phase involves the characterization and evaluation of the performance of the mill. Points of inefficiencies in the process are going to be identified in this phase.

Phase 3 – Constraint analysis:

The third phase involves the analysis of steam constraints and water constraints before extracting the data to build thermal composite curves. A set of guidelines are developed to provide a systematic approach to obtain theoretical energy targets. The guidelines also include a strategy to shift from theoretical targets to potential energy targets.

Phase 4 – Projects proposal:

In phase four, the existing heat exchanger network is evaluated and further inefficiencies are identified. The proposed projects are implemented into existing heat exchanger network to build new heat exchanger networks at different constraint levels. Economic analysis is done for each constraint level

The breakdown of the different phases in chronological order is presented in figure 3-1.

3.2 Breakdown of the phases

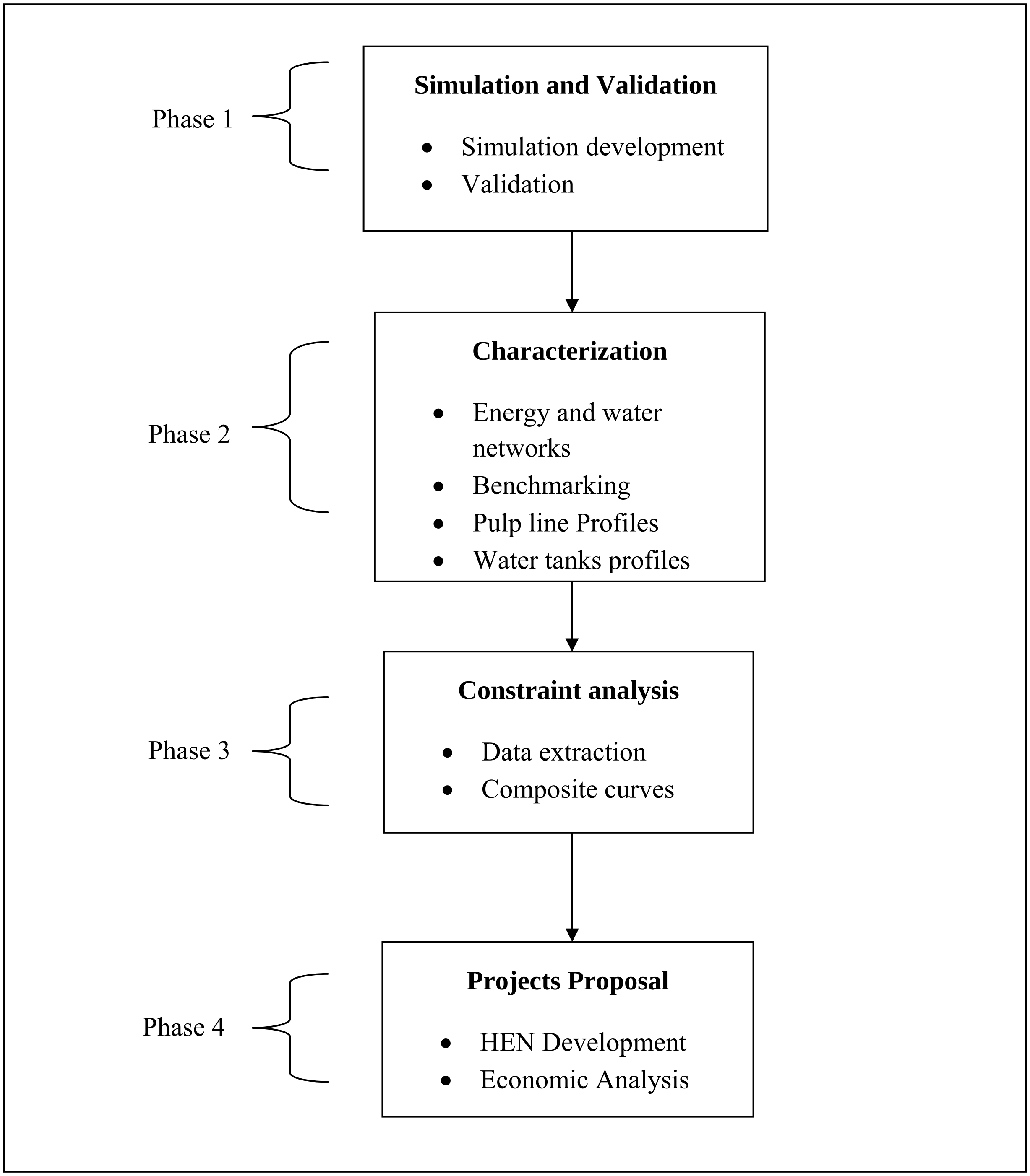


Figure 3-1: Breakdown of the methodology phases

3.3 Definition of the methods, techniques and tools

Simulation and validation

A base case simulation will be developed using a pulp and paper software called CADSIM plus®. The simulation will be based on data collected from the mill and will represent the actual heat and mass balances of the mill for winter conditions. The model will be presented to include the overall production of pulp, consumption of water and steam and fuel. To ensure the reliability of the simulated process, validating major results such as steam and water consumption with average mill data will be undertaken. In addition, the validation of the configuration will be done directly with mill staff.

Characterization:

In the characterization phase, water and steam networks are developed with the relevant stream information. Mass balance around these networks is done to account for all the steam and water production, utilization and post utilization. Key results from the mill will be benchmarked against the Canadian industry average. The results will include total water and energy consumption and effluent production. Key performance indicators such as the efficiency of boilers and the percentage of condensate return will be calculated and compared to the average Canadian industry. In addition, consistency and temperature profiles for the pulp line and water tanks are going to be plotted and examined. The outcome of applying all these tools is to locate energy inefficiencies such as non-isothermal mixing points.

Constraint analysis

The data required for the energy analysis will be extracted in two different ways: grassroot and retrofit. A study will be performed to find out the promising combination of extraction paths. The criteria used to evaluate the path will include minimum utility requirement, heat exchanger area, cost of heat exchanger and the effect on the energy bill. Guidelines will be developed to screen between the different types of extraction in the different process streams.

Using the relative sets of data from step three, thermal pinch analysis will be performed using Aspen Energy Analyzer®. The outcome of the thermal pinch analysis will provide the theoretical minimum heating and cooling required by the mill. This analysis will indicate the scope of the theoretical savings and the potential measures taken to reduce water and energy consumption.

Projects proposal:

Projects to increase the efficiency of the process will be proposed based on the constraint analysis results. The key ideas are going to revolve around increasing internal heat recovery, elimination of non-isothermal mixing points, and elimination of cross pinch transfers to increase the energy savings. The projects will be implemented in new heat exchangers at different constraint levels. The developed heat exchangers networks are going to be compared and evaluated based on the savings, capital cost, operating cost savings, and payback period.

CHAPTRE 4 SIMULATION MODEL, VALIDATION AND CHARACTERIZATION

4.1 Introduction

In any energy study, streams data consisting of mass and energy information is needed to perform energy analysis. The data could be used directly in the analysis or manipulated and used to build a dynamic mass and energy balance model. The model should be reliable and accurate with output and input data that resembles real life conditions of plants. Therefore the model needs to be validated in terms of output data and configuration. Once the validation is complete, a key step is to characterize and analyze the mill's energy and water performance and compare it to other existing mills.

This chapter will consist of three main parts; the first part will discuss the simulation model while the second part will discuss the validation process and finally the third part will discuss the characterization of the model/mill. In the first part the idea and the method behind developing a simulation model on CADSIM Plus® will be discussed in details. Followed by that, validation will be presented whereby the main focus is on the validation of configuration and output data. The main data to be validated is water and steam production and consumption. In addition, some other process parameters in the recovery loop will be addressed and validated.

The second part of this chapter will revolve around the idea of mill characterization. This will focus heavily on understanding the water and energy systems. This is a very important step in the methodology and will lead to the identification of inefficiencies in the process. By utilizing water and steam network diagrams, one can notice and identify the different constraints regarding steam consumption. This part of the methodology will be reflected in chapter 5. In addition, benchmarking of the mill's water and energy consumption as well as other key parameter indicators against the Canadian average will help to identify certain inefficient departments/units. This will help to filter out efficient departments and narrow down the scope of research. Finally, the pulp line temperature and consistency profiles will be presented as well as the hot water and warm water tanks temperature profiles. By examining these charts, non isothermal mixing in the pulp line and water tanks will be identified. Eliminating these inefficient mixing points will increase the energy savings thus making the mill more profitable.

4.2 Simulation Model

The simulation model has been developed on CADSIM Plus® software. It is a specialized program for the pulp and paper industry which creates a model with a mass and energy balance of the process. The model is created using different types of information from the mill. The configuration of the process is primarily based on process and instrumentation diagrams "P&ID's" as well as distributed control system "DCS" snap shots from the controls room. The input data for the system was based on P & ID's, DCS snap shots and excel files containing a long term average results for different streams in the process.

The mill consists of two continuous fibre lines operating simultaneously at the same time (Line A and Line B). The mill produces about 1600 oven dry tonne/day at a consistency of 90% or higher to be shipped to Canadian and international consumers. Line A was first built during the late 60's while line B was constructed during the late 70's. Both process lines have their own independent fibre lines and recovery loops. The chemical preparation department is shared between them. Each line has a power boiler and a recovery boiler to produce high pressure steam. In each line, high pressure steam is sent to a back pressure turbine to produce electricity and two lower levels of steam. High pressure steam is at 4300 kPa, 400 °C while medium pressure is at 1150 kPa, 202 °C and finally low pressure steam is at 450 kPa, 170 °C. The produced steam has a common header whereby steam is split between both lines based on the requirements of the process. Fresh water is heated through heat exchangers in the process to produce warm water and hot water. During the winter time, fresh water is at 2 °C while warm water is at 50 °C in line B and 57 °C in line A. Hot water is heated to temperatures of 65 °C in line B and 80 °C in line A. Bleaching chemicals are produced from chemical preparation department and sent to line A and line B. There is a continuous exchange of water, chemicals, liquor streams and steam between both lines. The amount of exchange varies depending on the requirement of each line at a specific point in time.

Both lines are almost identical in terms of configuration. In the cooking department, line A has an extra oxygen delignification step that is not present in line B. In addition line B has a two stage atmospheric diffuser while line A has a single stage pressure diffuser. In the evaporators department, line B consists of 6 effect evaporators and a concentrator while line A has a 5 effect evaporators and a cascade evaporator. Figure 4-1 represents the overall configuration of both lines.

Figure 4-1: Overall view of the model

Table 4-1 includes the major information of the model for both lines:

Table 4-1: Major Model information

Data	Line A	Line B
Chips Input (ODT/d)	1760	2023
Pulp Production (ODT/d)	724	864
Total yield (%)	41	43
Steam consumption (GJ/ADT)	20.9	23.9
Condensate returns (%)	56	51
Water consumption (m^3/ADT)	47.1	34.2
Effluent production (m^3/ADT)	48.5	35.8
Natural gas consumption (t/d)	5	30
Hog consumption (t/d)	0	3070
Electricity production (MW)	22.6	26.4

4.3 Validation of the model

The output/input data and the configuration of the developed model have to be validated thoroughly in order to have an exact representation of the mill's operating conditions. Working with a validated and reliable model will lead to a practical analysis with applicable results to the mill. Having many discrepancies between the model and mill's conditions will lead to a faulty analysis with impractical results. Therefore the validation step is a key point that should be done before starting the energy analysis of the mill.

The configuration of the model has to be verified directly with the mill staff. Depending only on P&ID's or DCS snapshots is not enough to validate the configuration. Major differences could exist between the P&ID's and the current situation of the mill because of the lack of continuous updates to old P&ID's. In addition, some of the information from the mill could be unreliable due to errors in measuring devices placed around the mill. Flow meters and consistency meters usually have a high percentage of error that could reach up to 50 % while temperature probes have an accuracy of around +/- 1 °C. In the case where there is a lack of long terms average values of a certain measurement and a discrepancy exist between the model and the mill value, one should suspect that the reason could be from the mill and not the model. In this case, more information is required and that could only be obtained by collaboration with the mill staff.

The input/output data of the model is validated with the mill staff and any other available information source such as DCS snapshots or long term average values on excel files. The best option is to always validate against long term average values if possible. There are three main categories of data to be validated in the mill:

1- Water production and consumption
2- Steam production and consumption
3- Other key parameters

Other key parameters will include flows of certain streams in different departments, temperatures of streams, production of pulp and consumption of white liquor. The validation results for the three categories are presented in the following subsection.

4.3.1 Water Validation - Line A

Water production through heat exchangers and consumption through the many consumers has been validated in table 4-2 against average long term values from the mill. There is a good fit between both sets of data with a difference of less than 5 %. This difference comes from the hot water usage in the pulp machine Slusher and the bleaching white water tank.

Table 4-2: Water validation - Line A

Department/ Consumer	Path	Model (L/min)	Mill (L/min)
Digester – Total		8993	9250
Digester - Cold blow cooler	WW to HW Tank	1777	1750
Digester - Flash Steam Condenser	WW to HW Tank	7215	7500
Washing – Total		1000	1000
Doctor board shower	WW to wire cleaning	1000	1000
Water Production - Make up Water	CW to WW tank	1018	2000
Bleaching – Total		13565	14400
Bleaching – Cold	CW to WW bleach Chest	300	300
Bleaching - doctor board shower	WW to wire cleaning	1000	1000
Brownstock dilution conveyer	HW	2000	2000
Do Showers	HW	4996	5000
Bleach White water chest	HW to bleach Chest	2268	3100
washer seal tank D1	HW as Make up	1000	1000
Contaminated condensate tank	HW as Make up	2001	2000
Other Consumers	HW	1	0
Evaporators – Total		8372	8500
Evaporators - Surface Condenser	CW to WW tank	8372	8500
Machine – Total		2619	2190
Machine	HW to Washer	429	0
Machine	CW to WW Chest	2190	2190
Total		35566	37340

4.3.2 Water Validation - Line B

Water production through heat exchangers and consumption through the many consumers has been validated in table 4-3 against average long term values from the mill. There is a good agreement between both sets of data with a difference of less than 1 %. This difference originates from the fresh water usage in the chemical preparation and the white water in bleaching showers.

Table 4-3: Water validation - Line B

Department/ Consumer	Path	Model (L/min)	Mill (L/min)
Digester – Total		11665	12200
Digester - Cold blow cooler	CW to WW Tank	3391	3700
Digester - Steam Condenser	WW to HW Tank	8274	8500
Washing – Total		3321	1900
Brownstock -doctor board shower	WW to wire cleaning	200	200
Brwonstock - Dilution conveyor	HW to conveyor	200	200
Brwonstock - Dilution conveyor	WW to conveyor	1505	1500
Brwonstock - Press Washer	HW	1416	0
Bleaching – Total		12468	9500
Bleaching - Showers	White water	4969	2000
Bleaching - doctor board shower	WW to wire cleaning	1000	1000
Bleaching Showers	HW to showers	6499	6500
Chemical Preparation		3073	7300
Chemical Preparation	CW	2978	6000
Chemical Preparation - R8	HW	96	100
Chemical Preparation - R8 Chiller	CW		1200
Evaporators		13207	13200
Evaporators - Surface Condenser	CW to WW tank	13207	13200
Machine		2023	2000
Machine - White Water Chest	CW to WW Chest	500	500
Machine – miscellaneous	WW	1523	1500
Total		45758	46100

4.3.3 Steam Validation - Line A

Steam production and consumption have been validated against average long term value and DCS snapshots. By examining table 4-4, one can notice that both the mill and model values have insignificant errors. This can be also said to the MP steam consumed in line A with error equals to 1 %. On the other hand, there is some suspicious discrepancy in the LP consumption and this is due to faulty measurements in the bleaching steam mixers, bleach heater and brown heater flow meters. The reasoning behind this error was obtained through numerous discussions with the mill personnel. The total error in LP steam is 20 %. This information is available in table 4-5.

Table 4-4: Steam Production - Line A

Department	Model (t/hr)	Mill (t/hr)
Total HP produced	193.8	194
HP Produced-RB1	193.8	194
HP Produced-PB2	0.0	0
HP- From B	0.0	0
HP-to Turbine	268.5	274
HP- to PRV`s	0.0	0
CD – to PRV’s	10.3	11.3
MP produced from turbine	77.6	80
LP produced from turbine	190.9	191

Table 4-5: Steam consumption - Line A

User	Model (t/hr)	Mill Data (t/hr)
Pulp-MP	10.3	10.15
02 Delignification-MP	5.6	6.4
P.M-MP	50.2	48.5
Evaporators-MP	1.8	1.8
Pulp-LP	13.8	12.75
Bleaching+ brown -LP	26.3	13
Bleach Heater - LP	12.1	3
P.M- Shower - LP	7.5	7.55
P.M- Lazy Shower- LP	24.0	23.95
Evaporator-LP	61.9	62.4
Recovery Boiler- AHX - LP	4.8	2.0
Deaerator-LP	23.5	20
Total MP - Line A	67.9	66.9
Total LP - Line A	173.9	144.7

4.3.4 Steam Validation - Line B

Steam production and consumption have been validated like line A. By examining the table 4-6, one can notice that both the mill and model values have small errors. MP steam consumed in line B has an error less than 1 %. Moreover, table 4-7 shows some suspicious results in the LP consumption with error of 10 %. Based on the mill personnel experience, the discrepancy is due to faulty measurements in the bleaching steam mixers, heater, and water condenser flow meters.

Table 4-6: Steam production - Line B

User	Model (t/hr)	Mill (t/hr)
Total HP produced	376.5	375
HP Produced-RB5	248.6	248
HP Produced-PB4	127.9	127
HP-to Turbine	292.2	286
HP-to line A	0.0	0
HP- to PRV`s	9.6	9
CD – to PRV’s	10.2	10
MP produced from turbine	62.1	65.9
LP produced from turbine	230.1	232.1

Table 4-7: Steam consumption - Line B

User	Model (t/hr)	Mill Data (t/hr)
Pulp-MP	29.03	29.95
Pulp Machine-MP	47.81	47.15
Evaporators MP	1.80	1.80
Recovery Boiler -MP	5.00	5.00
ClO2 Plant - MP	1.00	1.00
Digester-LP	27.24	30.50
Bleach Heater-LP	4.68	20.00
Bleaching-LP	26.12	43.00
P.M-LP	25.00	25.00
Evaporators-LP	77.72	78.60
Condensate Stripper - LP	17.00	17.00
Recovery Boiler- AHX - LP	5.93	4.00
Deaerator-LP	51.96	50.00
space heater + vent - LP	19.00	19.00
C.P-LP	6.25	6.85
Water Prod- Cond. - LP	3.81	0.00
Total MP - Line B	84.6	84.9
Total LP - Line B	264.7	294.0

4.3.5 Steam validation - Line A + B

Table 4-8 represents the total steam production and consumption in both lines. There is an agreement between the production and consumption data. The 30 t/hr that caused the 20% error in line A is consumed in line B. Once the adjustment factor is taken into account, the overall balance of steam production and consumption shows an error of less than 1%.

Table 4-8: Steam production and consumption - Line A and B

Department	Model (t/hr)	Mill (t/hr)
Total HP Produced	570.3	569
HP Produced-RB1	193.8	194
HP Produced-PB2	0.0	0
HP Produced-RB5	248.6	248
HP Produced-PB4	127.9	127
HP-to Turbine A	268.5	274
HP to turbine B	292.2	286
HP- to PRV's	9.6	9
Total consumption and production		
Total MP produced	152.5	151.8
Total LP produced	438.4	438.6
Total MP consumed	152.5	151.8
Total LP consumed	438.4	438.6
Overall Balance		
CD- TO PRV's (Steam)	20.5	21.3
Total Steam produced	570.3	569
Total Steam consumed	590.8	590.3

4.3.6 Other Parameters validation - Line A:

Table 4-9 contains many key parameters located at different streams in the process. The data is validated against snapshots from 2009 and 2010. There is a good fit between both sets of data except in natural gas consumption in RB1 and active alkali concentration in white liquor. Furthermore, the causticizing efficiency is a bit higher in the model than in the mill. All the parameters are presented in table 4-9:

Table 4-9: Other parameters validation - Line A

Department	Parameter	Mill - 2010	Mill - 2009	Model
Digester A	Ratio of organics to inorganic solids in WBL	1.7	-	1.60
	Dissolved solids concentration in WBL (%)	N/A	18	19
	Flow of WBL to evaporators (L/min)	5200	5237	5218
	Flow of white liquor (L/min)	N/A	2278	2244
Evaporators A	Temperature in effect 1 (°C)	120.5	120.1	119.4
	Temperature in effect 2 (°C)	120.5	119.8	120.2
	Temperature in effect 3 (°C)	110.9	109.5	109.5
	Temperature in effect 4 (°C)	93.5	91.5	91.5
	Temperature in effect 5 (°C)	78.5	75.1	78.8
	Temperature in effect 6 (°C)	58.5	52.2	58.4
	Dissolved solids concentration in SBL (%)	45.3	47	54.3
	Flow of SBL to recovery boiler (L/min)	1890	1171	1293
Steam Plant A	Make up Water (L/min)	1471		1204
	Natural Gas consumption in RB1 (m^3/hr)	N/A	0.05	0.12
	Air flow in RB1 (t/d)	N/A	8770	7750
	Excess 02 in RB1 (%)	N/A	3.7	2.30
	Natural gas consumption in PB2 (km^3/hr)	0	3.1	0.00
	HOG consumption in PB2	0	0	0.00
Recaust A	Caustisizing efficiency (%)	N/A	79.4	95.0
	Active Alkali g/l	N/A	104.9	294.2
	Sulphidity (%)	N/A	29.8	29.8
Bleaching	Bleach rate (t/d)	N/A	851	803

4.3.7 Other Parameters validation - Line B

Similar to line A parameters, line B parameters portray a good fit between the mill data and the model data. There is a discrepancy in the flow of strong black liquor from the evaporators to recovery boiler. This is due to the high dissolved solids concentration and low water content in the model. All the parameters are presented in table 4-10 below:

Table 4-10: Other parameters validation - Line B

Department	Parameter	Mill - 2010	Mill - 2009	Model
Digester B	Ratio of organics to inorganic solids in WBL	1.7	0	1.75
	Dissolved solids concentration in WBL (%)	N/A	17	19
	Flow of WBL to evaporators (L/min)	6000	5220	5676
	Flow of white liquor (L/min)	N/A	2405	2366
Evaporators B	Temperature in effect 1 (°C)	114.3	105.3	101.1
	Temperature in effect 2 (°C)	94.7	88	86
	Temperature in effect 3 (°C)	77.6	73.2	79.2
	Temperature in effect 4 (°C)	63.3	62.3	65.2
	Temperature in effect 5 (°C)	56.1	53.8	52.3
	Temperature in concentrator (°C)	116.8	108.6	115.2
	Dissolved solids concentration in SBL (%)	70	71	68
	Flow of SBL to recovery boiler (L/min)	1389	1327	935
Steam Plant B	Make up Water (L/min)	3888	0	3274
	Natural Gas consumption in RB5 (m^3/hr)	0.3	0.5	0.12
	Air flow in RB5 (t/d)	6960	7147	7247
	Excess 02 in RB5 (%)	N/A	2.36	2.30
	Natural gas consumption in PB2 (km^3/hr)	0.0543	0.6	0.12
	HOG consumption in PB4 (t/d)	756	490	499
Recaust B	Caustisizing efficiency (%)	N/A	81.8	95.0
	Active Alkali (g/l)	N/A	105.3	134.5
	Sulphidity (%)	N/A	29.1	29.2
Bleaching	Bleach rate (t/d)	N/A	951	949
LINE A+B	Total Machine Prod (adt/d)	N/A	1660	1764
	Total Digester Prod (adt/d)	N/A	1560	1949
	Total weak black liquor (L/min)	11048	10453	10895

4.4 Characterization of the model

The characterization of the validated model is an essential step of the analysis. There are two main outcomes from this step; the first is to enhance the knowledge and understanding of the process while the second outcome is the identification of inefficiencies in the process. In order to achieve these outcomes, the following tools are used:

1- Steam network: A diagram is developed consisting of all steam producers and consumers. In addition, steam is classified into direct injection and indirect injection. The condensate returning is identified as well. The diagram is supported by a table of stream information.

2- Water network: A diagram is developed consisting of water production and consumption cycle. In addition, heat exchangers required to heat the water are included. The effluents produced are identified as well. The diagram is supported by a table of stream information.

3- Benchmarking: The mill's steam consumption, water consumption, and effluent production are compared against the Canadian industry average. The departments operating below and above the average are highlighted and analyzed.

4- **Key Performance Indicators:** Certain parameters in the mill are compared to the average Canadian industry. These parameters or indicators include condensate return and boilers efficiency. This will help to indicate where general inefficiencies occur.

5- Pulp line profiles: The temperature and consistency profile along the pulp line have been plotted. The main idea is to use the diagrams to identify non isothermal mixing in the pulp line. In addition, it presents an excellent image of the consistencies across the line.

6- Water tanks profile: Hot water and warm water tanks flow vs. temperature profiles are plotted on a chart for input streams and output streams. Non isothermal mixing in tanks could be easily identified. The streams could be shifted between tanks to eliminate non isothermal mixing.

The use of these six tools will shorten the pathway of obtaining energy saving solutions and thus expanding the platform of ideas regarding the energy saving options in the mill. The results from these tools are discussed below.

4.4.1 Steam network

Line A steam network

In 4-2 below, steam production from the boilers and utilization in the different departments have been sketched. To complete the steam cycle, condensate or post utilization has been included in the diagram. In addition, fresh makeup water into the condensate tank has been highlighted. The red streams represent that exchange of HP, MP and LP between line A and line B. The corresponding tables for stream flows and temperatures are in appendix 1-1.

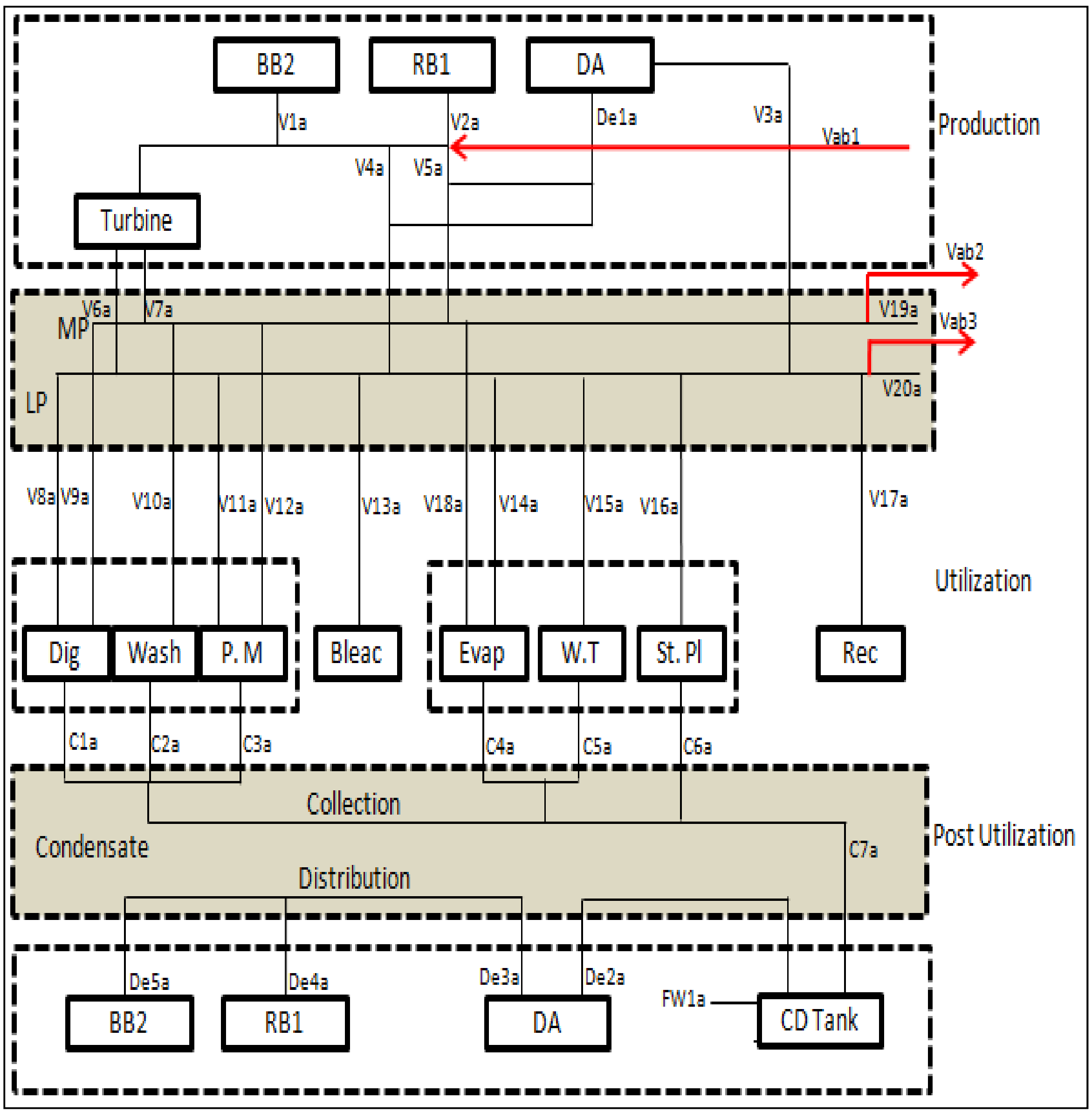


Figure 4-2: Steam network - Line A

The table below contains the steam balance around the network. The high pressure steam is produced from recovery boiler 1 and is sent to a back pressure turbine. Power boiler 2 is off at this time of the year. Water entering the boilers comes from the deaerator whereby it is heated with low pressure steam. Medium pressure steam and low pressure steam are produced and consumed in the different departments. The total consumption of medium pressure and low pressure steam is 242 t/hr. This steam is either used directly through different injection points in the digester, bleaching, and pulp machine departments or used indirectly through heat exchangers. Indirect steam use is more efficient since the condensate produced is sent back to the condensate collection tank and then the boilers. This will decrease the amount of energy needed to heat the water before the boilers. Information regarding the direct injected steam is presented in appendix 1-2. The condensate returns with a flow of 124 t/hr which is equivalent to 56 % of the total steam consumed. The other 44 % of steam is either directly injected or flashed and used at other points in the process. Medium pressure and low pressure steam is usually flashed to atmospheric conditions before being sent to the condensate collection tank. This occurs in the condensate of medium pressure and low pressure in the digesters, pulp machine, water production, steam plant, and evaporators departments. Directly injected steam flow is 97 t/hr while the flashed steam is 10 t/hr. By combining these values, the steam cycle is balanced. The breakdown of the steam flows in the cycle is presented in table 4-11.

Table 4-11: Steam balance - Line A

Definition	Value
Steam consumed (t/hr)	241.5
Condensate return (t/hr)	134 (56%)
Injected steam (t/hr)	97.1
Flashed steam	10.4
Injected steam + condensate+ flashed steam (t/hr)	241.5

Line B steam network

In similar manner to line A, steam production from the boilers and utilization in the different departments have been sketched on figure 4-3. Condensate returning as well as fresh makeup water into the condensate tank has been highlighted on the diagram. The red streams represent that exchange of HP, MP and LP between line A and line B. The corresponding tables for stream flows and temperatures are in appendix 1-1.

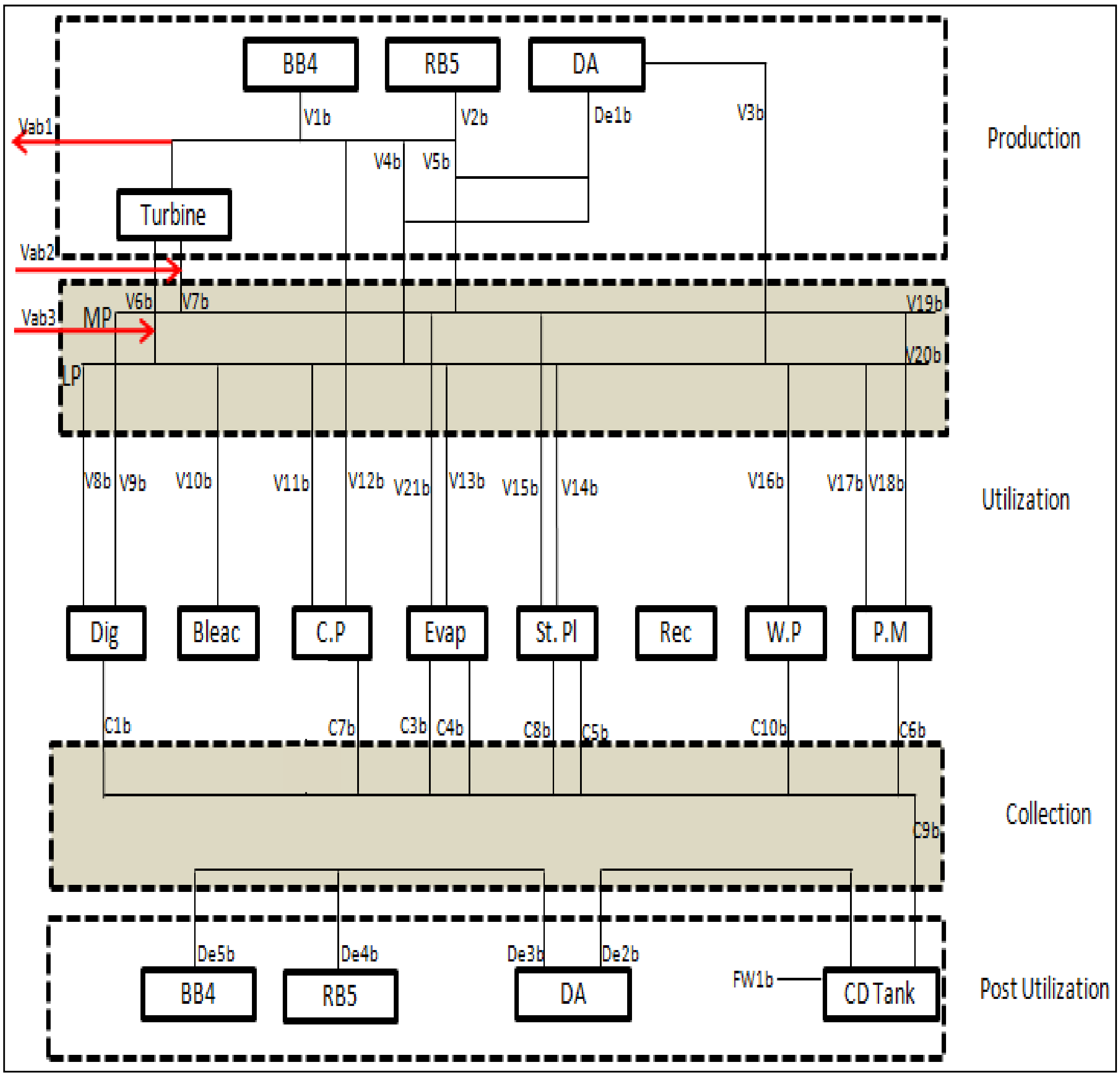


Figure 4-3: Steam network - Line B

The table below contains the steam balance around the network. The high pressure steam is produced from recovery boiler 5 and power boiler 4 and is sent to a back pressure turbine. Medium pressure steam and low pressure steam are produced and consumed in the different departments. The total consumption of medium pressure and low pressure steam is 347 t/hr. This steam is either used directly through different injection points in the digester, bleaching, and pulp machine departments or used indirectly through heat exchangers all over the mill. Direct steam is injected in different streams and equipment such as steaming vessel, water tanks or in the pulp line. The steam injection is a constraint in many cases whereby it cannot be replaced with other hot streams due to operational difficulty. More information regarding the role of steam as a constraint or non constraint will be discussed in chapter 5. In addition, Information regarding the directly injected steam is presented in appendix 1-2. The condensate returns with a flow of 175 t/hr which is equivalent to 51 % of the total steam consumed. The other 49 % of steam is either directly injected or flashed and used at other points in the process.. The directly injected steam flow is 167 t/hr while the flashed steam is 5 t/hr. By combining these values, the steam cycle is balanced. The breakdown of the steam flows in the cycle is presented in the table below:

By comparing line A and line B steam consumption, it is obvious that line A condensate return rates are higher than line B by 5%. A larger amount of directly injected steam is lost in the steaming vessel and the deaerator. In the latter case, a large amount of fresh makeup water enters the condensate tank which needs to be heated in the deaerator using directly injected low pressure steam. This cyclic heating of makeup water and loosing condensate has a negative effect on the efficiency of the cycle. The breakdown of the steam flows in the cycle is presented in table 4-12.

Table 4-12: Steam balance - Line B

Definition	Value
Steam consumed (t/hr)	346.6
Condensate returns (t/hr)	174.9 (51%)
Injected steam (t/hr)	166.5
Flashed condensate (t/hr)	5.1
Condensate returns + Injected steam + Flashed steam (t/hr)	346.6

4.4.2 Water network

Line A water network

Figure 4-4 below represents the water network in line A. Fresh water (FW) is heated through different heat exchangers across the mill to produce warm water (WW) and hot water (HW). Some of these heat exchangers consume steam while other depends on internal heat recovery to transfer the required energy load. The list of heat exchangers and streams information is presented in appendix 1-3. More information regarding the heat exchanger network is presented in chapter 6. Water is then utilized in the process and sent to the sewer. In addition water from other streams such as chemical solutions in bleaching are included to close the balance.

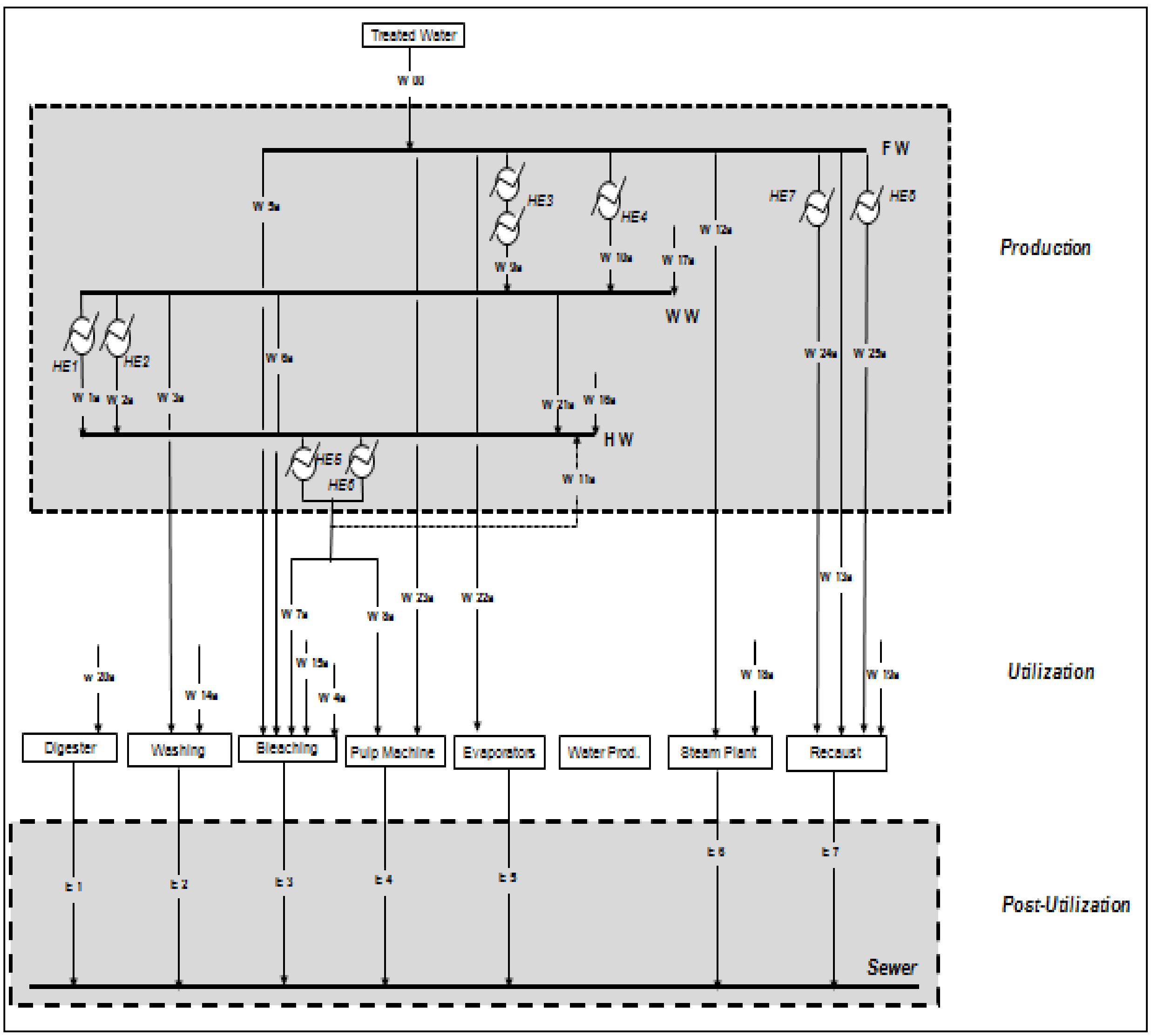


Figure 4-4: Water network - Line A

The water balance for fresh water, warm water, and hot water for line A is presented in table 4-13 below. The difference between the production and consumption of water ranges from 0% to 2%. This is due to the fact that volumetric flow rate of water streams in the simulation model is affected by the temperature. Therefore a small difference is apparent in each type of water.

Table 4-13: Breakdown of water balance – Line A

Source	Total (m3/hr)	m3/hr	%
Fresh water entering	1311		
Fresh water used	1297		
Error		14	1.1%
Warm Water Produced	788		
Warm Water Used	791		
Error		-3	-0.3%
Hot water produced	780		
Hot Water Used	762		
Error		18	2.3%

The overall balance of water for line A is presented in table 4-14. The total fresh water consumed plus water from other sources as well as water from line B equals the total effluents produced. The small difference in the balance is due to the effect of temperature on volumetric flow rates.

Table 4-14: Overall water balance - Line A

Overall Balance	Total (m3/hr)	m3/hr	%
FW used	1311		
Water from other resources	285		
Water from B to A	333		
Total Fresh Water	1929		
Error		14	0.7 %
Total Effluents	1915		

Line B water network

Figure 4-5 below represents the water network in line A. Fresh water (FW) is heated through different heat exchangers across the mill to produce warm water (WW) and hot water (HW). Some of these heat exchangers consume steam while others depend on internal heat recovery to transfer the required energy load. The list of heat exchangers and stream information is presented in the appendix 1-3. More information regarding the heat exchanger network is presented in chapter 6. Water is then utilized in the process and sent to the sewer. In addition water from other streams such as chemical solutions in bleaching are included to close the balance.

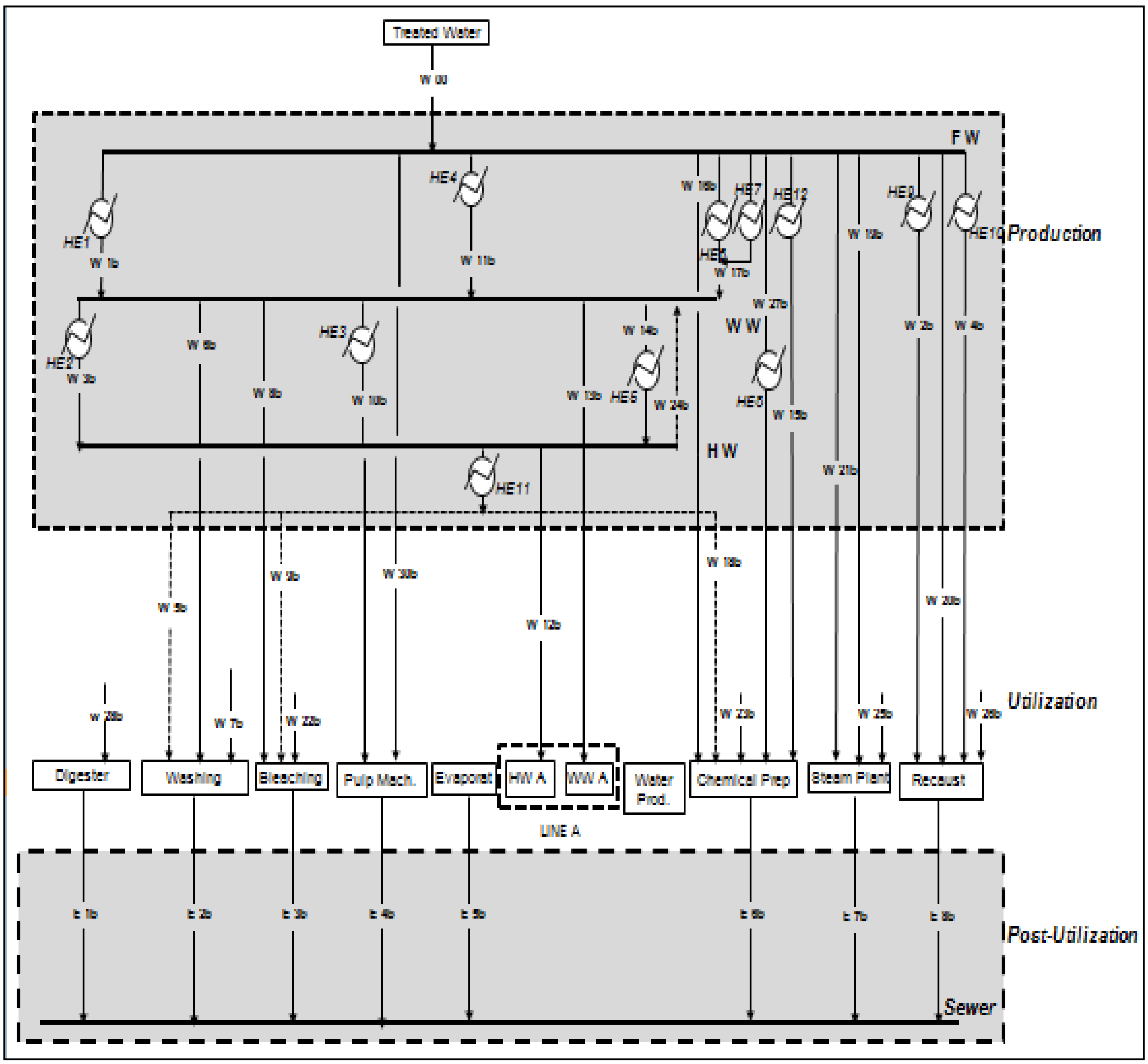


Figure 4-5: Water network - Line B

The water balance for fresh water, warm water, and hot water for line B is presented in table 4-15. The difference between the production and consumption of water ranges from 0% to 1%. This is due to the fact that volumetric flow rate is affected by the temperature of the stream. Therefore a small difference is apparent for each type of water.

Table 4-15: Breakdown of water consumption - Line B

Source	Total (m3/hr)	m3/hr	%
Fresh water entering	1686		
Fresh water used	1698		
Error		-12	-0.7%
Warm Water Produced	1069		
Warm Water Used	1073		
Error		-4	-0.3%
Hot water produced	600		
Hot Water Used	601		
Error		-1	-0.1%

The overall balance of water for line B is presented in table 4-16. The total fresh water consumed plus water from other sources equals the total effluents produced. In addition the water that is sent to line A is considered as an effluent to close the balance for line B. The small difference in the balance is due to the effect of temperature on volumetric flow rates.

Table 4-16: Overall water balance - Line B

Overall Balance	Total (m3/hr)	m3/hr	%
Fresh water used	1686		
Water from other resources	251		
Total fresh water	1937		
Error		16	0.9 %
Effluents	1611		
Water going to A	309		
Total Effluents	1921		

4.4.3 Benchmarking

Thermal consumption

Thermal consumption for the main departments in line A and B is presented in figure 4-6. Line A and B are treated either as separate mills or as an integrated mill. The thermal consumption of steam in the steam plant is not included in the benchmarking due to the lack of data regarding the use of steam in the deaerator. The benchmarking data is obtained by surveying a number of Canadian mills and finding the median, modern and 25^{th} percentile mills [9].

It is evident that the mill performance cannot be compared to the modern mills. In the digesting department the consumption of thermal energy is higher than the median and the 25^{th} percentile. It can be seen that line B consumption is higher than line A; steam consumption in the steaming vessel and liquor heaters is much higher than line A. In the pulp machine, Line A steam consumption is much higher than line B. This is due to high amount of steam needed to heat fresh water in the white water chest. The total consumption for the dried pulp is below the median but above the 25^{th} percentile while the total consumption for the dried pulp is below the median and the 25^{th} percentile. This indicates that steam savings could be obtained in the digester and pulp machine. Additional savings could be possible in all the departments to match modern or 25^{th} percentile mills.

Figure 4-6: Benchmarking - Thermal consumption

Water Consumption

The water consumption for the main departments of line A and B is presented in figure 4-7. Line A and B are treated either as separate mills or as an integrated mill. The benchmarking data is obtained by surveying a number of Canadian mills that were designed in the 1980s and finding the average for these mills[23].

The total water consumption for line A is higher than the 1980's design while line B is lower. This is because line B was built during the late 1970's while line A was built during the 1960's. This integrated mill has total water consumption below the 1980's design. By focusing on specific departments, one can notice that the bleaching A, pulp machine A, evaporators are high consumers of water. Line B departments have lower water consumption than line A and 1980's design. In the recovery boiler departments, the water composes of fresh makeup water to the boilers and water into stripper condensate. Stripper condensate only exists in line A and therefore the water consumption in line A is higher. Finally, the makeup water in both lines is more than the 1980's design and this is because the condensate that returns in the process is around 55% while in newer designs the number is much higher.

Figure 4-7: Benchmarking - Water consumption

Effluent Production

The effluent production for the departments of line A and B is presented in figure 4-8. Line A and B are treated either as separate mills or as an integrated mill. The benchmarking data is obtained by surveying a number of Canadian mills that have 1990's design and finding the average for these mills[24, 25].

The total effluent production for line A is higher than the 1990's design while line B is lower. This is evident due to the fact that water in consumption in line A is higher than B and therefore more effluents are produced. The total water consumption for the integrated mill is above the 1990's design. By focusing on specific departments, one can notice that bleaching A and B, pulp machine A, evaporators are high producers of effluents. Line B departments have a relatively lower production of effluents. The reason for the high effluents in bleaching is the high intake of water. In evaporators line A has condensate stripper that used a lot of fresh water. The combined water is then sent to the sewers and is not used thus increasing the production of effluents. There is a possibility of redesigning the mill and lowering the effluents production by increasing water reutilization and therefore reducing the total water consumption.

Figure 4-8: Benchmarking - Effluent production

4.4.4 Key Performance Indicators

Key performance indicators were calculated for both lines. The calculated values are then compared to the Canadian industry average[2]. The results are presented in table 4-17. The condensate return for both lines is below the average Canadian industry. This indicates that there is potential to reduce the amount of steam injection and increase the condensate return. Both recovery boilers efficiency is below the average Canadian industry as well. More work could be done on the configuration of boilers to enhance its efficiency. This work will not be done in this project. The steam consumption is higher than the Canadian industry average in both lines; this is also seen in the benchmarking results. The discrepancy between the two values is due to the change in units and the addition of all steam consumers. Finally, the water consumption in both lines is much lower than the Canadian industry average and therefore one can suspect that the savings in energy consumption are going to be more significant than the savings in water consumption.

Table 4-17: Key Performance Indicators

Key Performance Indicator	Unit / Line	Mill	Canadian Average
Condensate Return (%)	Line A	56	60
	Line B	51	
Recovery Boilers Efficiency (%)	RB1	57.5	60
	RB5	51.2	
Biomass Boiler Efficiency (%)	BB2	N/A	80
	BB4	88.4	
Steam Consumption (GJ/adt)	LINE A	20.89	18.5
	LINE B	23.46	
Water Consumption (m3/adt)	LINE A	47.1	75.0
	LINE B	34.2	

4.4.5 Pulp line profiles

Temperature profile – Line A

The temperature profile of pulp line A is presented in figure 4-9. The temperature of the pulp can be seen at different sections of the process. In the cooking section, there is a temperature peak due to heating and cooking of pulp. As the pulp enters the washing section it maintains a constant temperature until the oxygen delignification step. The temperature increases to 90 °C due to direct medium pressure steam injection. This is the first non isothermal mixing point appearing in the pulp line. In the bleaching section, the pulp undergoes heat and cooling due to washing steps and steam injection. Five major non isothermal mixing points are apparent in the pulp line. These points are going to be addressed and eliminated in chapter 5 and 6. In the machine section, pulp maintains a steady temperature before it is heated to 90 °C at three stages: steam showers and indirect steam injection in the dryer.

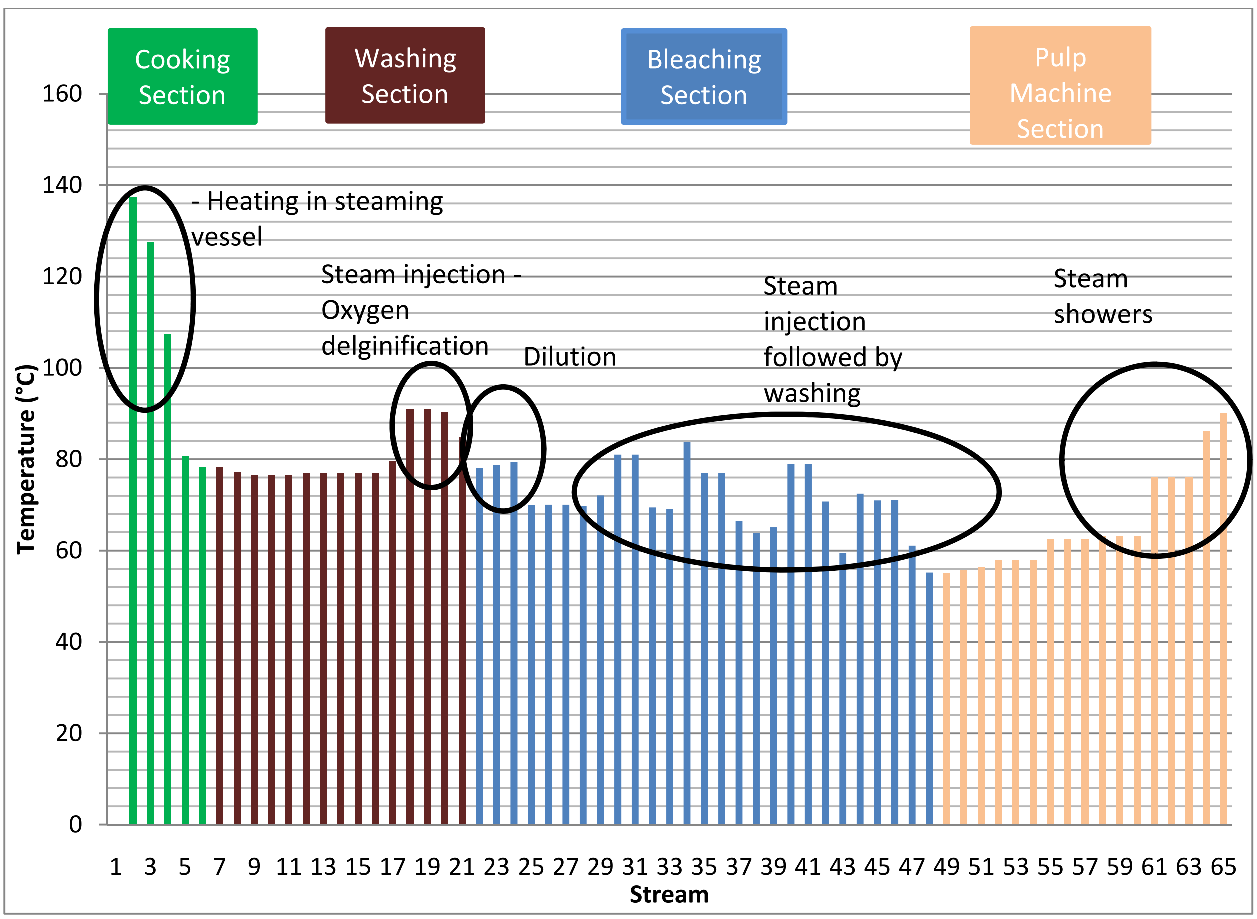


Figure 4-9: Temperature profile - Line A

Consistency profile – Line A

The consistency profile is a great tool to analyze the consistency levels across the pulp line. In the cooking section, the pulp enters at a high consistency before it is diluted with cooking liquor in the digester high pressure feeder. The consistency changes significantly in the washing section due to the dilution and thickening in the drum washer and Decker washer. The consistency levels changes from 1% to 12 % in the washing section. Bleaching consistency profile shows a similar pattern to the temperature profile in the bleaching section. This is due to the systematic counter current thickening and dilution in the washers. In the last stage, pulp is pressed, dried and formed into sheets of pulp with a consistency of 90% before it is shipped to consumers. The consistency profile is presented in figure 4-10.

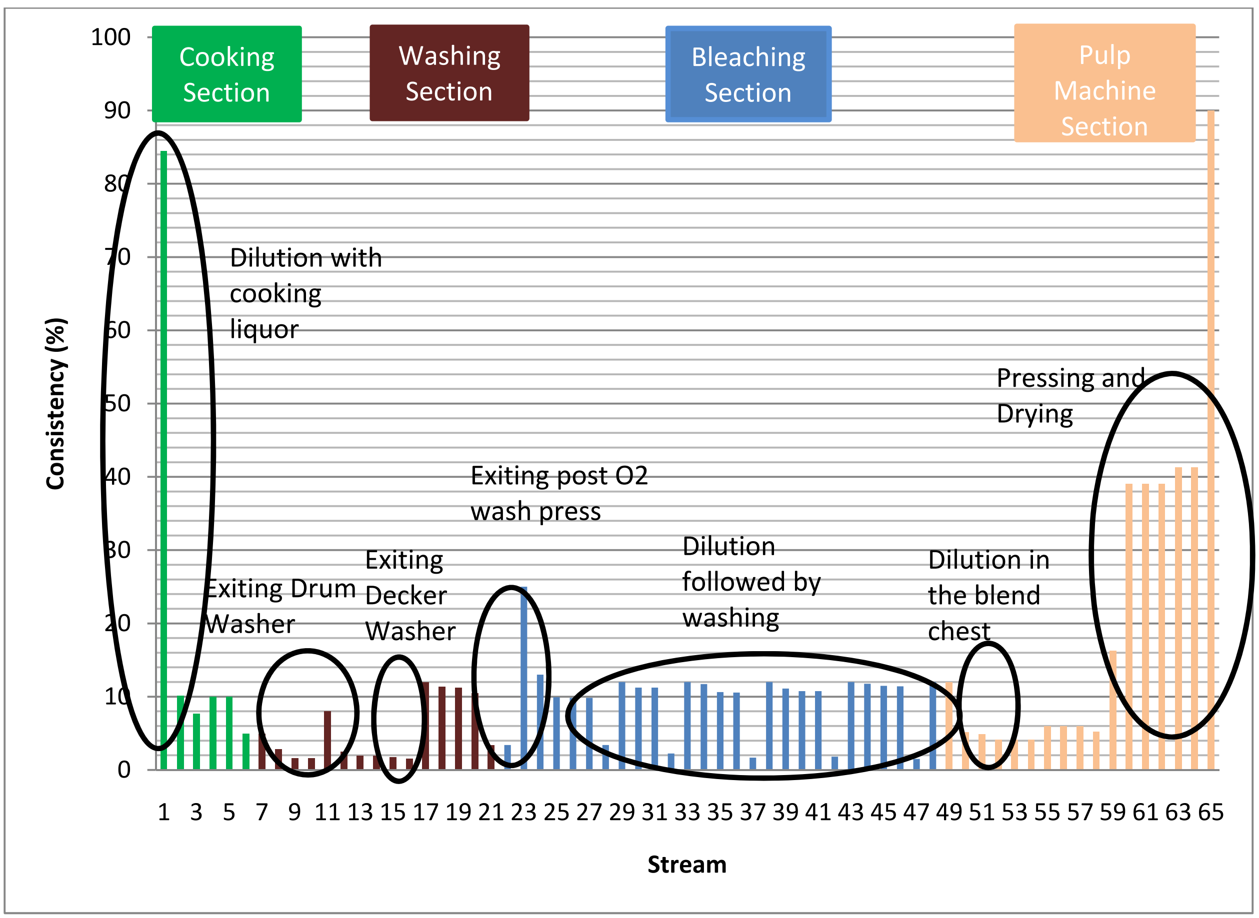


Figure 4-10: Consistency profile - Line A

Temperature profile – Line B

In the cooking section, the temperature is at its maximum levels due to heating and cooking of pulp. The temperature drops significantly after the pulp exits the atmospheric diffuser. As the pulp enters the washing section it maintains a constant temperature until the last dilution stage. This is one of the major non isothermal mixing points in the pulp line. In the bleaching section, the pulp undergoes heat and cooling due to washing steps and steam injection. Four major non isothermal mixing points are apparent in the bleaching pulp line. These points are going to be addressed and eliminated in chapter 5 and 6. In the machine section, pulp maintains a steady temperature before it is heated to 90 °C by water showers, steam showers and indirect steam injection in the dryer. The temperature profile of pulp line B is presented in figure 4-11.

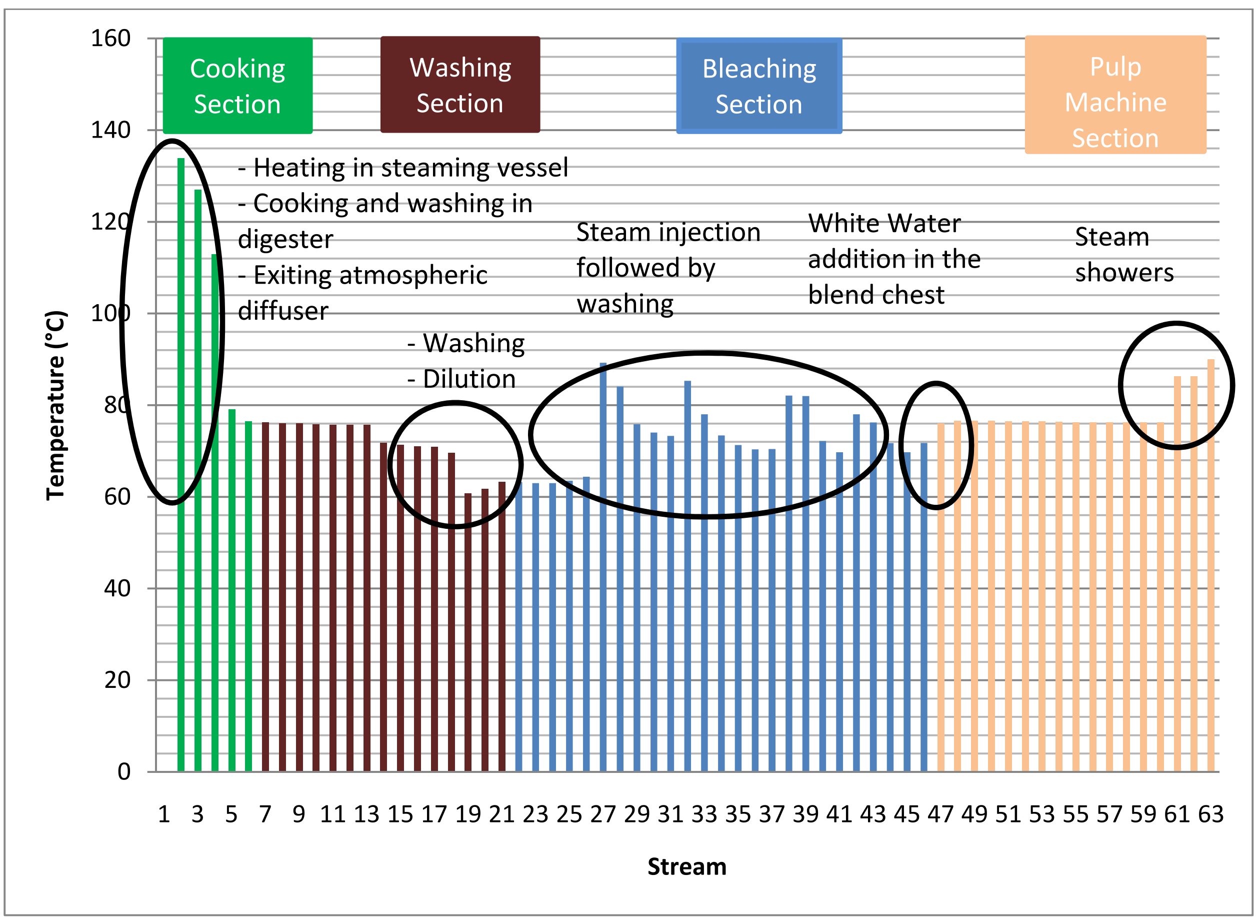


Figure 4-11: Temperature profile - Line B

Consistency profile – Line B

Line B has a similar consistency profile to line A whereby the pulp is diluted in the cooking section from a consistency of 85% to a consistency of 8%. In the washing section, the pulp undergoes dilution to bring down the consistency to 1%. The pulp consistency peaks at 30% after the Drum washer. This is followed by dilution to bring the consistency down to 5%. In the bleach section, multiple counter current thickening and dilution in the washers results in peaks and troughs resulting in a similar profile to the temperature profile. In the last stage, pulp is pressed, dried and formed into sheets of pulp with a consistency of 95%. The consistency profile is presented in figure 4-12.

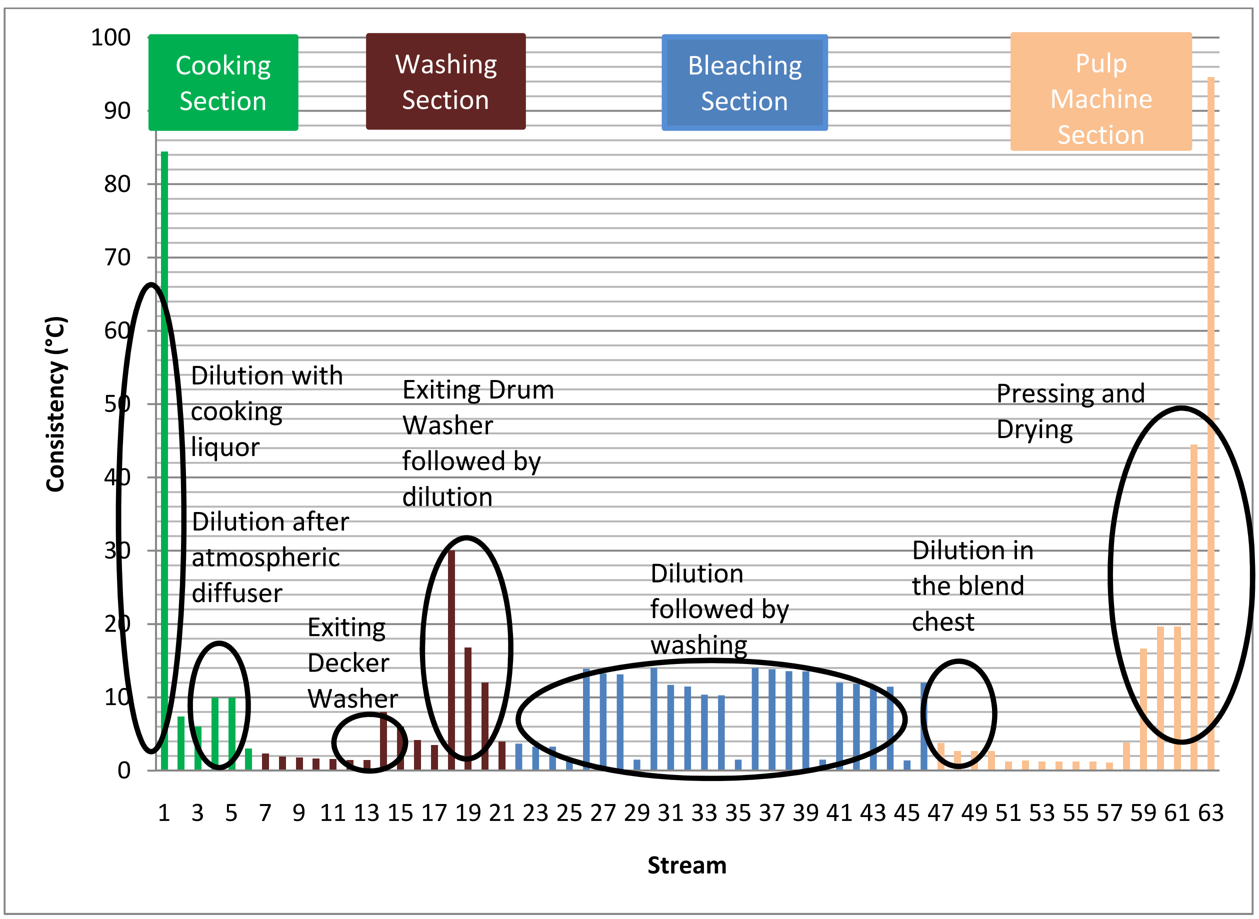


Figure 4-12: Consistency profile - Line B

4.4.6 Water tanks profiles

Warm water and hot water tanks - Line A

The streams entering the warm and hot water tanks are plotted based on their relative temperature and flows. The total output and the tank temperature are highlighted by the dotted line. Based on the profiles presented in figure 4-13, non isothermal mixing occurs in the warm water tank due to the use of cold fresh water to maintain the level in the tank. In the hot water tank, water from two different sources enters the tank with 20 °C difference. This results in non isothermal mixing in the hot water tank. The required temperature at different consumers is highlighted in the box on the right hand side of figure 4-13. The water network for line A is presented in chapter 5.

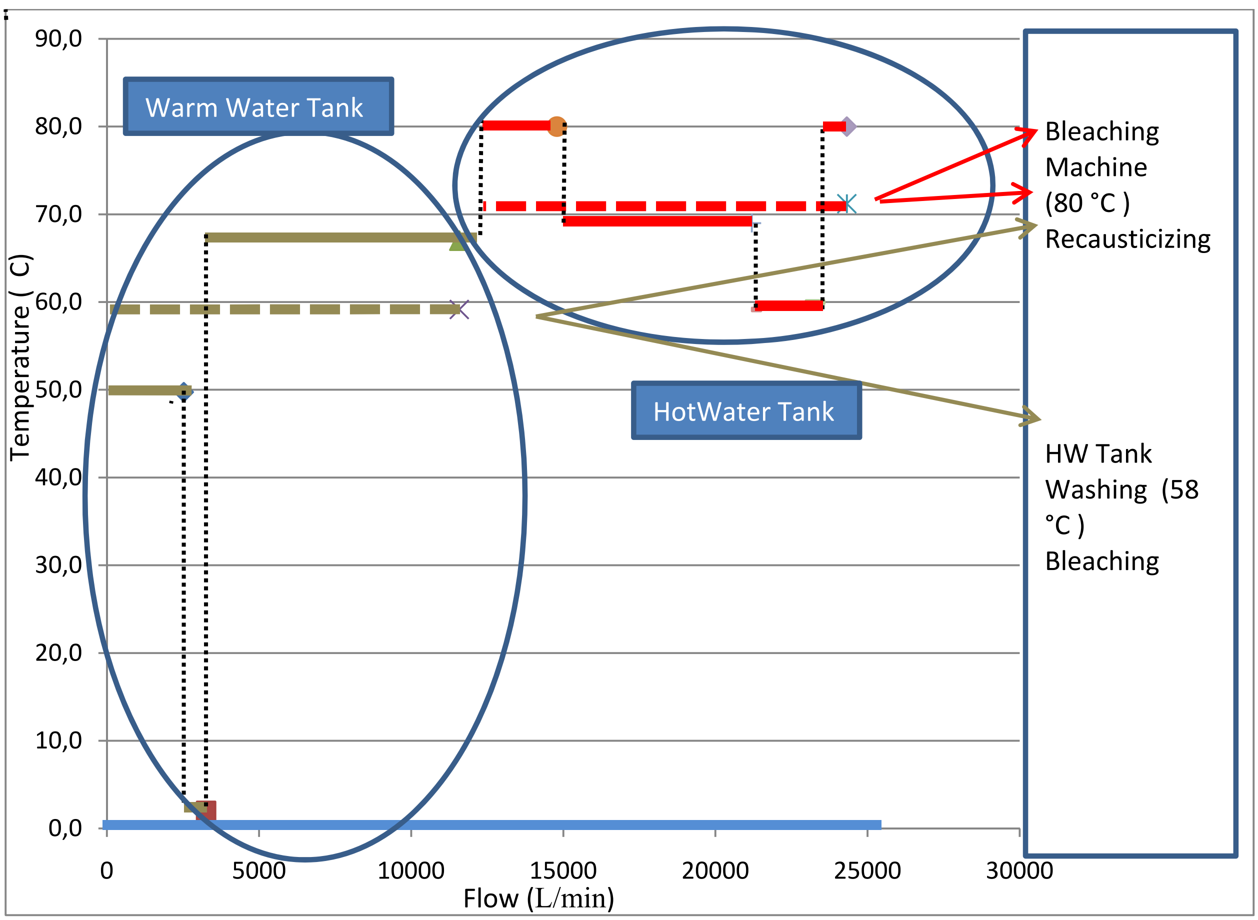


Figure 4-13: Warm water and Hot water tanks profile - Line A

Warm water and hot water tanks – line B

Warm water and hot water tanks for line B were plotted in the same manner. Based on the profiles presented in figure 4-14, non isothermal mixing is not as apparent as in line A warm and hot water tanks but it exists mainly in the warm water tank. More efficient allocation of streams could be done in the tanks thus resulting in a more efficient use of energy when streams are being heated. The required temperature at different consumers is highlighted in the box on the right hand side of the figure. The water network for line A is presented in chapter 5.

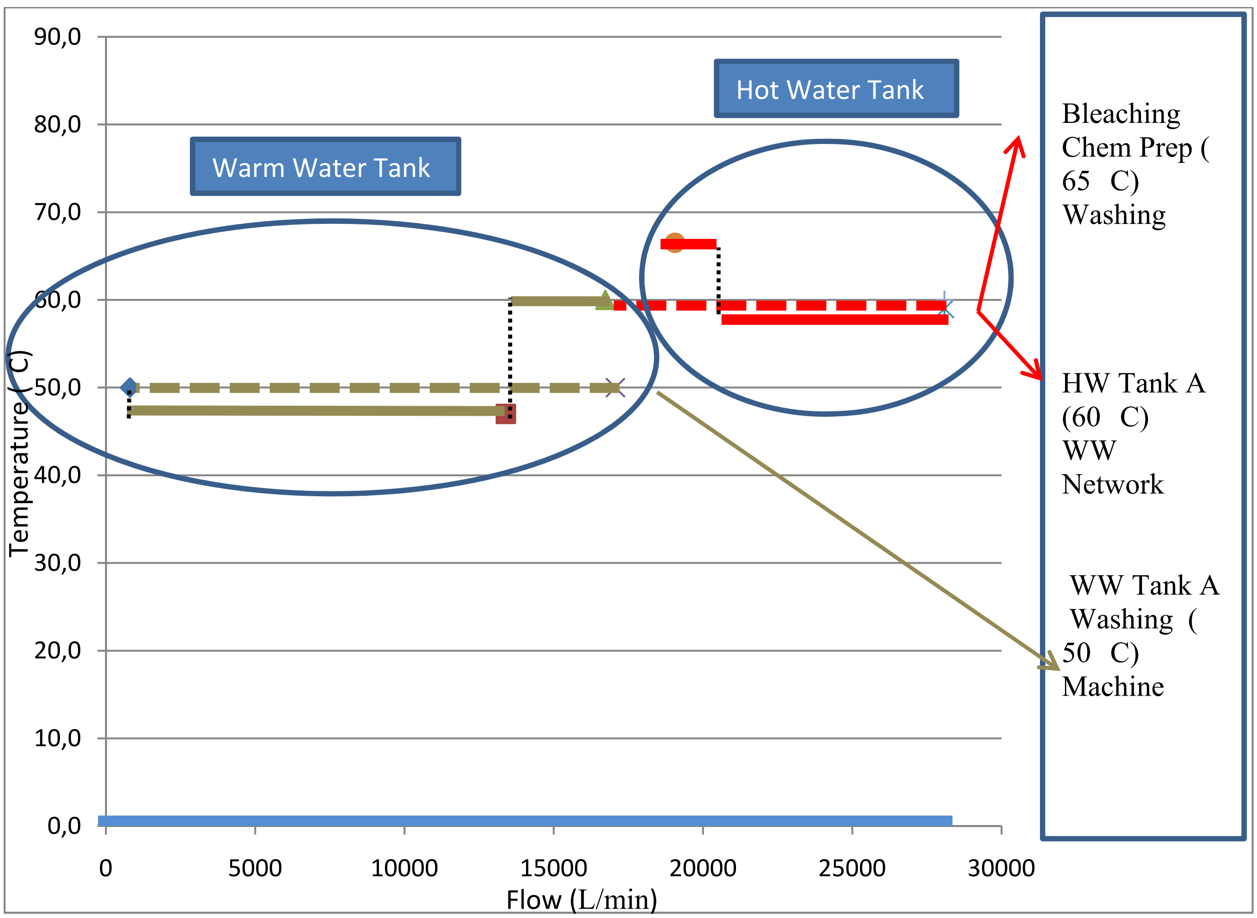


Figure 4-14: Warm water and hot water tanks profile - Line B

CHAPTRE 5 GUIDELINES TO CONSTRAINT ANALYSIS IN A KRAFT MILL

5.1 Introduction

Existing mills which have been built in the past three decades require periodical improvement to maintain profitability and advantage over newer mills. The failure of doing so results in higher operating costs due to the build-up of inefficiencies in the process. A set of techniques that were developed in the past four decades to overcome inefficient process designs are combined under a term called process integration (PI). PI is the use of techniques to improve energy and water efficiency of a process and reduce green house gas emissions by developing optimized designs. The most common techniques in PI are exergy analysis, thermal pinch analysis, and mathematical programming. The thermal pinch analysis is the most widely used technique today.

Process integration techniques lead to improvements by changing process conditions and operability, optimization of utility systems and water system, and debottlenecking of crucial sections in the process[13].

Pinch analysis is a concept that was added to the process integration pool in the late 1970's. It identifies efficient ways to recover internal heat from process streams and lowers dependence on utilities which leads to significant economical savings and higher return on investment for new or existing plants. Pinch analysis has been used in many industries including the petrochemicals, oil refining, and pulp and paper. The degree of savings depends on the efficiency of the existing design and economic aspects such as capital cost and return on investment. The scope of average energy saving for the industries mentioned above is between 15 % - 30% [14]. For the purpose of this project, a pinch analysis is applied.

In order to apply the thermal pinch analysis, key data such as process streams undergoing thermal change need to be extracted. The level of data extraction strongly depends on the existing constraints for a process, equipments and certain streams. Identifying the constraints is essential to develop a concrete and consistent approach to analyze and represent these constraints and reflect them as distinct levels of improvements. In this work, a set of guidelines is developed for the constraint analysis in Kraft processes.

5.2 Constraint analysis guidelines

The objective of analyzing constraints and developing guidelines is to provide a systematic approach to identify energy saving projects based on a realistic theoretical energy target. Applying the proposed guidelines assists users to correctly identify energy saving projects in a reasonably short period of time without depending on previous experience. Performing constraint analysis results in a range of projects with different levels of investments. The analysis and guideline presented in this study is general yet detailed enough to be applicable and extendable to any Kraft mill.

The focus of this chapter is on understanding the role of utility systems in the mill and the method for extracting and presenting the steam and water streams in the thermal composite curves. Utility steam could be considered as an irreplaceable heat source for the process interpreted as "steam constraint" or as a replaceable heat source reflected as "steam non-constraint". The water network could be presented as with a fixed constrained configuration known as "retrofit" or with a flexible non constrained configuration known as "grassroot". Therefore, different approaches of stream data extraction are explored based on the constraints of each heat transfer point. The effect of different constraints is portrayed on the final energy target of the mill. Energy targets and potential energy saving projects are obtained at different constraint levels and compared against their economic viability. The different constraint levels will be discussed in depth in this chapter as well as chapter six.

The guideline consists of nice main steps which are discussed in the following order:

1. In-depth knowledge of the process
2. Identifying and screening the heat transfer points
3. Categorizing the heat transfer points
4. Listing possible energy savings projects
5. Identifying possibilities for grassroot and retrofit representation
6. Definition of scenarios
7. Building the composite curves
8. Addition of non isothermal mixing elimination projects
9. Building the refined composite curves

Steps 8 and 9 are discussed in the results section. Figure 5-1 highlights the major steps in the methodology for constraint analysis and building the heat exchanger network.

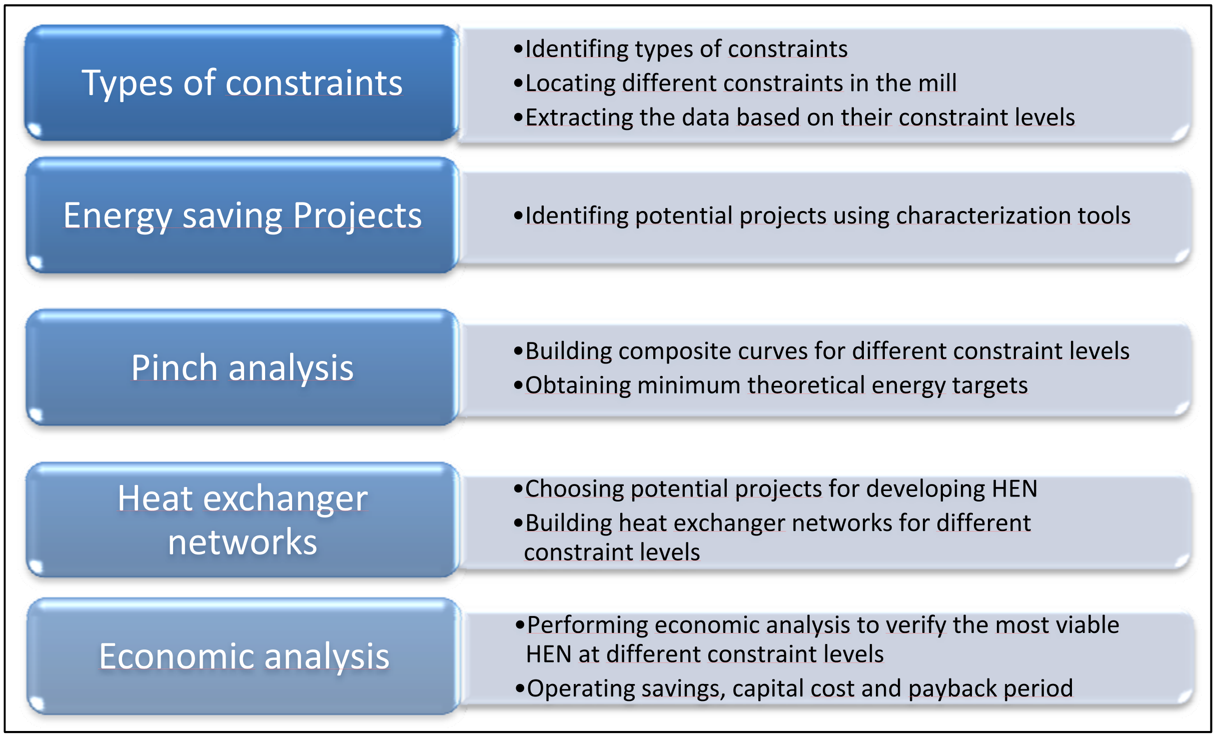


Figure 5-1: Overview of constraint analysis and heat exchanger network methodology

5.2.1 In-depth knowledge of the process

In this specific study, the goal is to analyze the energy and the water systems and propose modifications to reduce the overall operating cost. Generally, the key step to any energy study is acquiring an excellent knowledge of the process. Therefore the first step of the guidelines is to have depth knowledge of the process. To better master this task, the following steps which are explained in chapter four are recommended:

1- Drawing steam network by understanding where and how the utility steam is being consumed. The focus should be on the equipment or process demands that could result in steam constraints.
2- Drawing water networks by understanding where water and effluents are produced. Effluents are an excellent heat source to be utilized for internal heat recovery

3- Utilizing pulp line screening tools such as temperature and consistency profile in order to identify non isothermal mixing points in pup line. This could shed some light on the elimination of non isothermal mixing points and reduction of steam consumption.
4- Utilizing tanks screening tools to identify non isothermal mixing in water tanks. The elimination of these points will result in large steam savings in the water heating cycle.

5.2.2 Identifying and screening the heat transfer points

For any energy study, a set of reliable data is required to establish realistic and viable energy saving projects. Therefore process streams and key energy transfer points should be identified and tabulated. Data extraction of key streams varies based on constraints of energy utilization in a process and possible prospect projects. Changes in enthalpy and temperature are the two criteria for extracting heat transfer points with considerable energy content.

Generally a transfer point with enthalpy change above 300 kW and/or the temperature difference of 10 °C is considered to be a promising point in the analysis. These points could be dilution point in pulp lines, or steam injection point or water mixing in tanks. However, heat transfer points with $\Delta H > 300$kw and $\Delta T < 10$ °C, which are large flow rate with low energy content, are not subjected for extraction.

5.2.3 Categorizing the type of heat transfer point

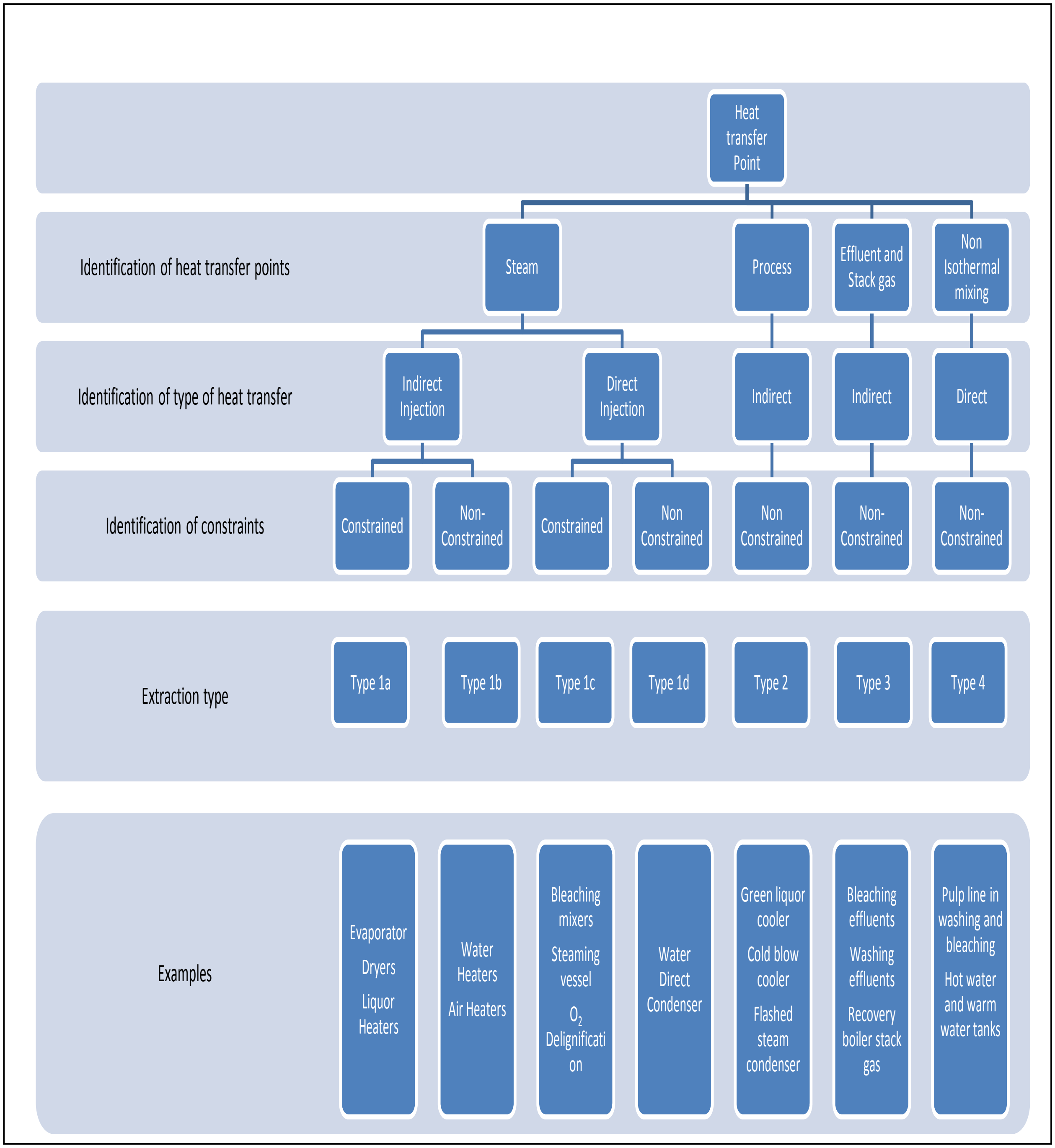
There are different ways of extracting, and representing data are mostly experience based and dictated by the objective of the energy analysis. For an example, heat exchange between streams could be indirect i.e. in a heat exchanger or direct through mixing points. Each type of energy exchange has a different representation of the extracted data with distinctive impact on the analysis and prospect projects. The following types of heat transfer points occur in a Kraft process and are addressed in this guideline

1. Steam injection points:
    a. Constrained indirect
    b. Non-Constrained indirect

   c. Constrained direct
   d. Non-Constrained direct
2. Indirect Heat transfer between different process streams
3. Energy in effluents and stack gases
4. Non isothermal mixing in water tanks and pulp line

Figure 5-2 shows the type of heat transfer points with the involved constraint type. Each path on the schematic leads to an efficient screening procedure for the constraints analysis. The first level is involved with the identification of heat transfer points by focusing mainly on their location and nature in the process. In this level, there are 4 points of heat transfer ; steam, process, effluents and gases, and non-isothermal mixing in water and pulp lines. The second level of the diagram discusses the identification of the type of heat transfer process. The focus in this step is to determine whether the heat is transferred through direct mixing or indirectly through heat exchangers. The third level is the most important step whereby heat transfer points are classified based on their flexibility and tendency to be changed. The essence of this level is to determine how soft or hard the constraints are and to evaluate their exact necessity for the process. An important question that needs to be answered at this point is whether the elimination or redesign of a heat transfer point in the process could hinder the operation or affect the quality of the final product; which in here is pulp. The outcome of this level is to reduce or eliminate the steam consumption at different heat transfer point in the process. This steam, if replaced by another heat source, would result in reduction of total steam consumption and therefore lowering the operating costs in the mill. A detailed look at the different types of stream data extraction is presented in the following section. Figure 5-2 is the diagram used in categorizing process constraints types:

Figure 5-2: Categorization of constraint type



The seven different types of extraction are discussed in depth with emphasis on the major discrepancies between the extraction techniques.

1a. Constrained indirect steam injection

Indirect steam injection is considered as a constraint in the following equipments/units:

- Evaporators
- Dryers
- Liquor heaters at high temperature

The steam demand in these units is critical and cannot be replaced with any other sources. By presenting the steam demand as a constraint, it is ensured that the theoretical energy target obtained by pinch analysis is closer to the potential energy target based on possible energy savings project. The following example in figure 5-3 shows the cooking liquor heater (Digester department) and the way that data is extracted.

⇩

Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Hot	202	185	Steam	S.ΔT.Cp
Cold	154	187	Liquor	L.ΔT.Cp

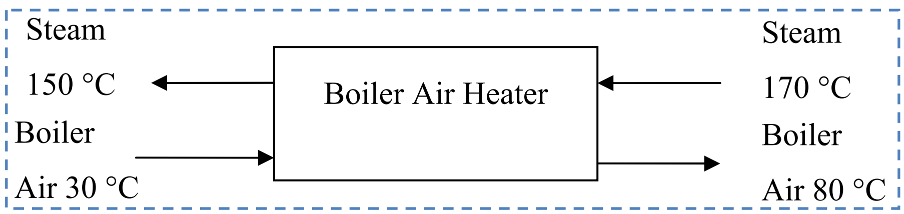
⇩

Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Cold	201	202	Steam	S.ΔT.Cp

Figure 5-3: Constrained Indirect steam Injection - 1a

<u>1b. Non constrained indirect steam injection</u>

Steam is used indirectly to heat liquor and water in the process. Since the hot stream is always "utility" steam, it never mixes with the process and returns to the boilers as condensate. In this type of extraction, steam consumption acts as a heating demand without constraints which is assessed and represented as the energy demand in the cold stream. In other words, the opportunity of replacing steam at this point and reutilizing internal heat recovery is reflected in the type of extraction. This type of heat transfer can be found in the air heaters and water heaters and the following example in figure 5-4 illustrates boiler air heater in steam plant.



⇩

Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Hot	170	150	Steam	S.ΔT.Cp
Cold	30	80	Air	A.ΔT.Cp

⇩

Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Cold	30	80	Air	A.ΔT.Cp

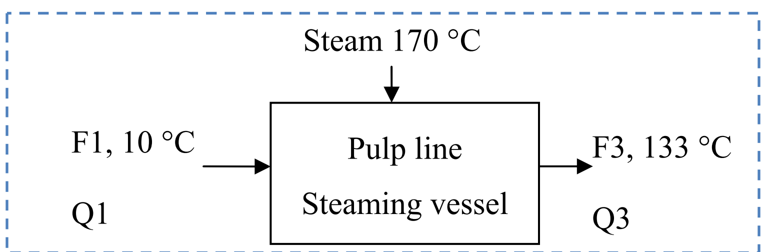
Figure 5-4: Non constrained indirect steam injection -1b

<u>1c. Constrained direct steam injection:</u>

Direct steam injection is usually utilized to increase the temperature of the process (pulp, water or liquor) to the required temperature. Pulp streams tend to exist at high consistencies in different parts of the process such as washing and bleaching. Pumping the high consistency pulp through a heat exchanger would increase the chances of clogging and fouling. Due to the impracticality of using an indirect heat source, direct steam injection is used even though it results in less efficient heating and loss of condensate. This is a case where steam is a constraint and cannot be replaced with another source. This concept of steam constraint is also applied to other parts of Kraft process whereby steam consumption is inevitable. Figure 5-5 shows the mill steaming vessel where pulp is heated with the low pressure steam from 10 °C to 133 °C. This is a critical step before cooking the chips. Since pre-steaming the chips can only be done with steam, the heating demand cannot be supplied by another source in the process and is represented as energy at steam temperature levels. In order to represent it as heating demand and not cooling, the Tin and Tout are switched and is considered as part of the cold streams. In order for the steam injection with constraint to be part of minimum heating requirement of the process, the temperature level Tin should be changed to the saturation temperature or 1 °C less than the steam temperature.

The direct steam injection for the pulp line occurs in the following equipments/units and the extraction should be based on the steam injection with constraint. The complete list of all direct and indirect steam injection constraints in a Kraft process are presented in appendix 2-1.

- Steaming vessel
- Bleaching mixers
- Steam showers
- Desuperheating
- Deaerator
- Tanks



Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Hot	170	133	Steam	S.ΔT.Cp
Cold	10	133	F3	

Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Cold	169	170	Steam	S.ΔT.Cp

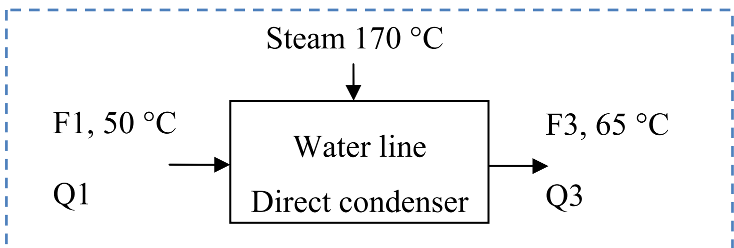
OR

Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Cold	169	170	Steam	Q3-Q1

Figure 5-5: Constrained direct steam injection – 1c

1d. Non constrained direct steam injection

Utilizing utility steam to heat water or liquor through a direct injection point is considered as steam consumption with no constraints and therefore it is subjected to be replaced with another hot stream. The direct steam injection for the water line occurs in the direct condenser whereby the extraction should be based on the technique presented below. In the following example in figure 5-6, warm water is being heated with low pressure steam from 50 °C to 65 °C. Water can be easily pumped through heat exchangers to utilize other energy sources in the process at lower temperatures. Therefore the cold stream is chosen to represent the energy demand for water. The heat demand could be directly calculated by subtracting the energy flow in F3 from the energy flow in F1 or using the steam energy flow.



Stream	T_{in} (°C)	T_{out} (°C)	Flow(t/d)	Q (MW)
Hot	170	65	Steam	S.ΔT.Cp
Cold	50	65	F3	F3.ΔT.Cp

Stream	T_{in} (°C)	T_{out} (°C)	Flow(t/d)	Q (MW)
Cold	50	65	F3	F3.ΔT.Cp

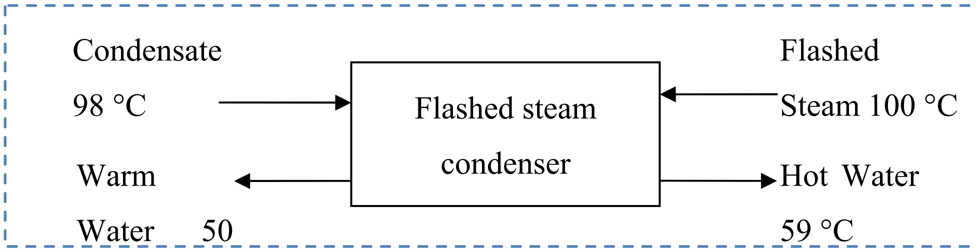
OR

Stream	T_{in} (°C)	T_{out} (°C)	Flow(t/d)	Q (MW)
Cold	50	65	F3	Q3-Q1

Figure 5-6: Non constrained direct steam injection 1d

2. Indirect Heat transfer between different process streams

In this simple case of data extraction a hot stream in the process cools down by releasing the energy to another stream through an indirect contact heat exchanger. Both the hot stream and cold streams are considered in the composite curves. Even if the flashed steam is used, it should be considered since it is part of the process rather than the utility system. This type is presented in figure 5-7 and is found in the cold blow coolers, glycol heaters, and green liquor coolers.

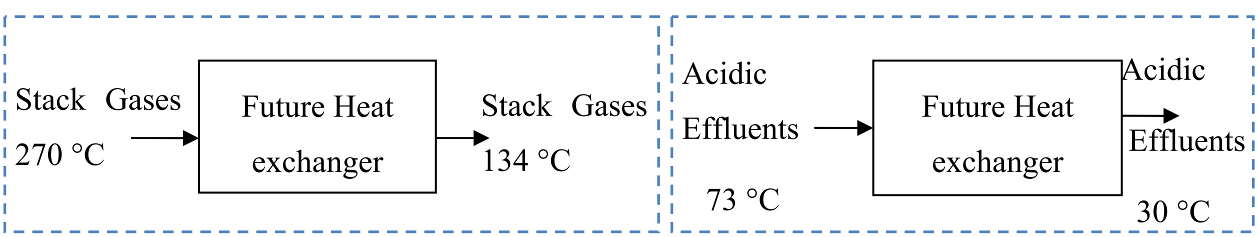


Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Hot	100	98	Steam	S.ΔT.Cp
Cold	50	59	WW	WW.ΔT.Cp

Figure 5-7: Indirect Heat transfer between different process streams - 2

3. <u>Energy in effluents and gases</u>

Effluents and flue gases contain a significant amount of energy. This energy can be reutilized through internal heat recovery to minimize the total heating requirement of the process and reduce the steam consumption of the process. Special attention should be paid to the lower temperature limits of effluents and gases due to condensation of corrosive components in these streams at low temperature. For an example, stack gases from recovery boilers are acidic and therefore the minimum temperature of the gas should always be kept above the dew temperature of the acid gas such as SO_2. Otherwise, acidic components would condensate and severely corrode the heat exchangers especially during start-ups and shut downs. The following two examples in figure 5-8 represent how the data extraction is carried out.



Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Hot	270	134	Gas	G.ΔT.Cp

Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Hot	73	30	Effluent	E.ΔT.Cp

Figure 5-8: Energy in effluents and gases - 3

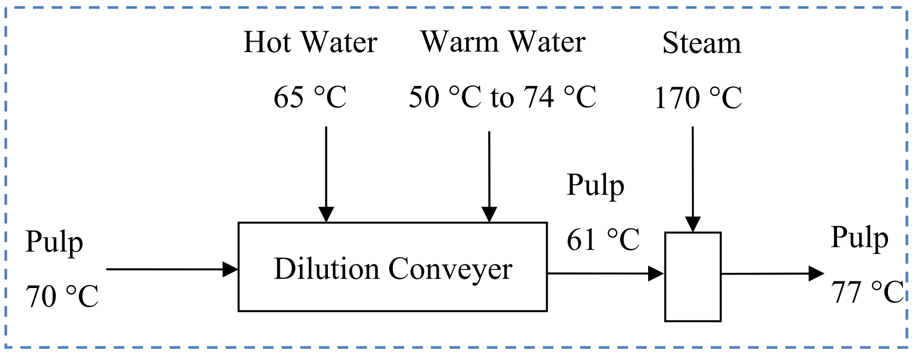
Effluents and gases are basically released to sewer or environment without recovering their energy. In order to identify them a detailed analysis of the water and steam network is required. Then, the effluent and gases streams with significant heat loads are considered as a viable heat source. In some cases, dirty effluents need to be filtered and cleaned before using them in a heat exchanger. Generally, the effluents and stack gases with the highest energy content are found in the bleaching department and the steam plant.

4. Non isothermal mixing – Pulp line and water tanks

The pulp dilution with water at low temperatures normally results in non isothermal mixing points and could lead to major inefficiencies. These inefficiencies occur when the temperature difference between the streams is large enough to result in a Criss-cross or a cross pinch heat transfer. Although Criss-cross heat transfer is not a pinch violation, but it occurs when a hot stream at a considerably higher temperatures and energy content is used to satisfy a cold demand at low temperature. This mixing leads to a higher heating and cooling demands in the process. The modifying approach to solve this problem is to assume that there are literally no constraints in the current configuration for heat exchanger network and water tanks. Non isothermal mixing is mainly noticed in the pulp line and water tanks. The procedure to solve these problems is discussed below:

a. Pulp line

Monitoring the temperature profile of pulp line reveals that due to certain temperature demand in bleaching towers, it is needed to be heated prior to bleaching stage. The main idea for data extraction is to eliminate the current non isothermal mixing points by proposing a new way to heat the pulp line in order to reduce the direct steam injection in the bleaching stage. This is done by heating water or liquor streams before mixing them with pulp. Then these new streams that need to be heated to lower temperature levels than steam will replace the constraint for steam demand (type 1c). This is found in the bleaching steam mixers or dilution conveyers. This reduction in steam consumption is reflected in the composite curves whereby a new minimum heating and cooling requirements are achieved. In figure 5-9, Water is heated before being fed in to the dilution conveyer in washing department using other hot streams available in the process. This project significantly reduces the steam consumption in the bleaching steam mixers. The target temperature of the water and the pulp vary depending on the location of non isothermal mixing points in the process. Therefore each constraint is addressed specifically depending on the process requirements. The data extraction for one specific project is presented in figure 5-9.



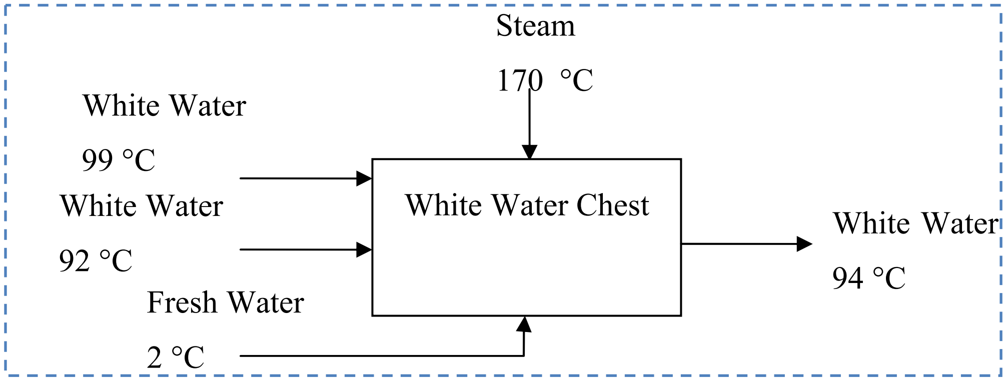
Stream	T_{in} (°C)	T_{out} (°C)	Flow (t/d)	Q (MW)
Hot	169	170	Steam	S.ΔT.Cp
Cold	50	74	WW	WW.ΔT.Cp

Figure 5-9: Non isothermal mixing in pulp line - 4

In this example, steam saving is achieved by choosing a target temperature of 74 °C and maintaining the temperature of hot water. Then this project is applied in the simulation of the mill and the portion of the steam that could not be saved is represented in the composite curve based on type 1c “constrained direct steam injection”.

b. <u>Water tanks</u>

Data extraction for hot and warm water tanks follows the same extraction procedure as pulp line mixing points. The only difference is the freedom of using heat exchanger in the absence of high consistency pulp that cannot be practically heated in heat exchangers. . Therefore Low consistency streams such as water and white water could be easily integrated into a new heat exchanger network before being used in the process. The following example in figure 5-10 represents the extraction technique used.



Stream	T_{in}(°C)	T_{out}(°C)	Flow (t/d)	Q (MW)
Cold	169	170	Steam	S.ΔT.Cp
Cold	2	94	FW	FW.ΔT.Cp

Figure 5-10: Non isothermal mixing in water tanks - 4

Increasing the temperature of fresh water to 94 °C results in steam saving. The portion of steam that could not be saved in this project is represented in the composite curves as type 1c data extraction. This type of constraints and projects occur in the hot water, warm water tanks and other types of water or liquor tanks.

5.2.4 Listing possible energy savings projects

Applying the constraint analysis in the overall process results in detecting the prospect possible projects. These projects usually occur around water-steam heaters or air-steam heaters. At this stage of the analysis, the identified projects involve heating a cold stream without knowing the location of the hot stream. The hot stream will be identified at the stage of building the heat exchanger network which is explained in chapter 6. In addition, most of the non-isothermal mixing elimination projects or type 4 constraints are not known at this point and they are presented when refining the thermal composite curves of the overall process. The initial potential projects for line A and line B are presented below in table 5-2 and 2-3 respectively.

Table 5-1: Initial energy saving projects - line A

Project name	Steam Saving (MW)
Bleach heater	7.37
Brown heater	3.24
Boiler Air Heater	3.16
Total	13.77

Table 5-2: Initial energy saving projects - line B

Project name	Steam Saving (MW)
Bleach heater	2.82
Direct Condenser	4.40
Boiler Air Heater	7.59
Total	14.82

The first three projects in line A and line B involve replacing steam by another heat source in order to heat water or air. In the three cases, LP steam is consumed in an indirect heat exchanger except for the direct condenser whereby steam is directly injected and condensed to heat water. The schematic of these projects could be found in the chapter 6. Schematics of all potential energy saving projects are presented in the heat exchanger network chapter. It is important to note that at the stage of constraint analysis the “potential projects” are identified. Evaluating the feasibility of these projects should be investigated by developing a heat exchanger network to designate the proper hot stream to satisfy the energy demand of each project. This issue is discussed in detail in chapter 6.

5.2.5 Identifying possibilities for grassroot and retrofit representations

In order to represent the stream information in the composite curves, data should be extracted with the correct heat load, temperature levels and configuration. Two different approaches for data extraction were studied to quantify the extent of energy saving with its associated economic aspects. Therefore the overall process is evaluated from the stand point of grassroot and retrofit data extraction.

In grassroot approach heat transfer points are represented without considering any configuration constraints in the process. The objective of this representation is to reach minimum target energy demand in the process regardless of the extra capital cost involved in building a new heat exchanger design. However, in retrofit approach modifications are bounded to the current existing design and heat transfer points are restricted to the current heat exchanger network configuration. Thus the objective in retrofit approach is to minimize the capital cost and any changes in the existing design while trying to reach the minimum energy target.

In the retrofit mode of data extraction, stream information is extracted based on the current configuration. The design constraints such as temperature and flow demands of the process are considered fixed and are carried to the future design. Therefore, most of the heat exchangers or mixing points will remain in place with the replacing the heat source or sink entering the heat exchanger. Normally the proposed retrofit projects are economically more viable and the capital costs are kept low with the modification limited to upgrading existing exchangers, purchasing reasonable number new of exchangers and upgrading the piping system.

In the grassroot mode of data extraction, the current configuration is not considered as a constraint while extracting data for energy analysis. The only constraint in this type of extraction includes the target flow and temperature demand of the process. Therefore, heat exchangers and tanks in the current design are not considered in the composite curves. The overall heat exchanger network will drastically change and could result in a higher energy savings possibility and higher capital costs. The following example presented in figure 5-11 and 5-12 illustrates the difference between the two types of data extraction:

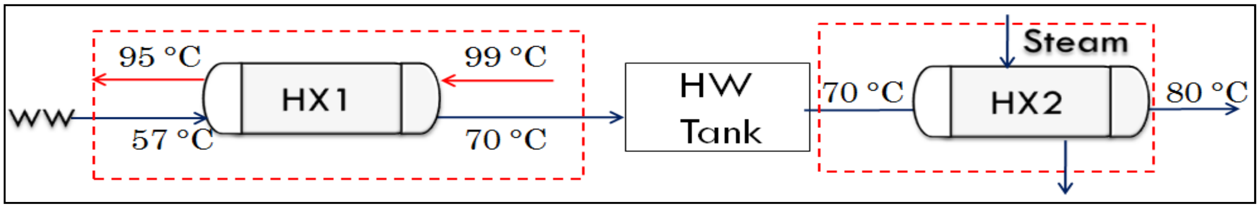


Figure 5-11: Retrofit approach

Retrofit case:

The current heat exchanger and tank configuration are considered as process constraint. The input and output of the hot water tank should be respected at all times. Any change could only occur to the hot streams entering the heat exchangers. The cold streams are fixed in terms of temperature and flow rate. Due to this restricted yet cost effective targeting technique, the extraction is represented in the following form:

HX1: $ColdStream = 57\ ℃ \rightarrow 70\ ℃\ , \text{Hot Stream} = 99℃ \rightarrow 95\ ℃$

HX2: $ColdStream = 70\ ℃ \rightarrow 80\ ℃$

Based on the retrofit extraction method, it is possible to achieve savings of 0.52 GJ/ADT by eliminating the cross pinch heat exchange at 65 °C and reutilizing stack gas instead of LP utility steam.

Grassroot Case:

In the grassroot representation, the input and output flow and temperature of the system are considered as constraints. The current configuration is not respected and is subject to change. Any heat exchanger, tank, or a hot stream could be replaced and changed with a simpler system such as the one presented in 5-12.

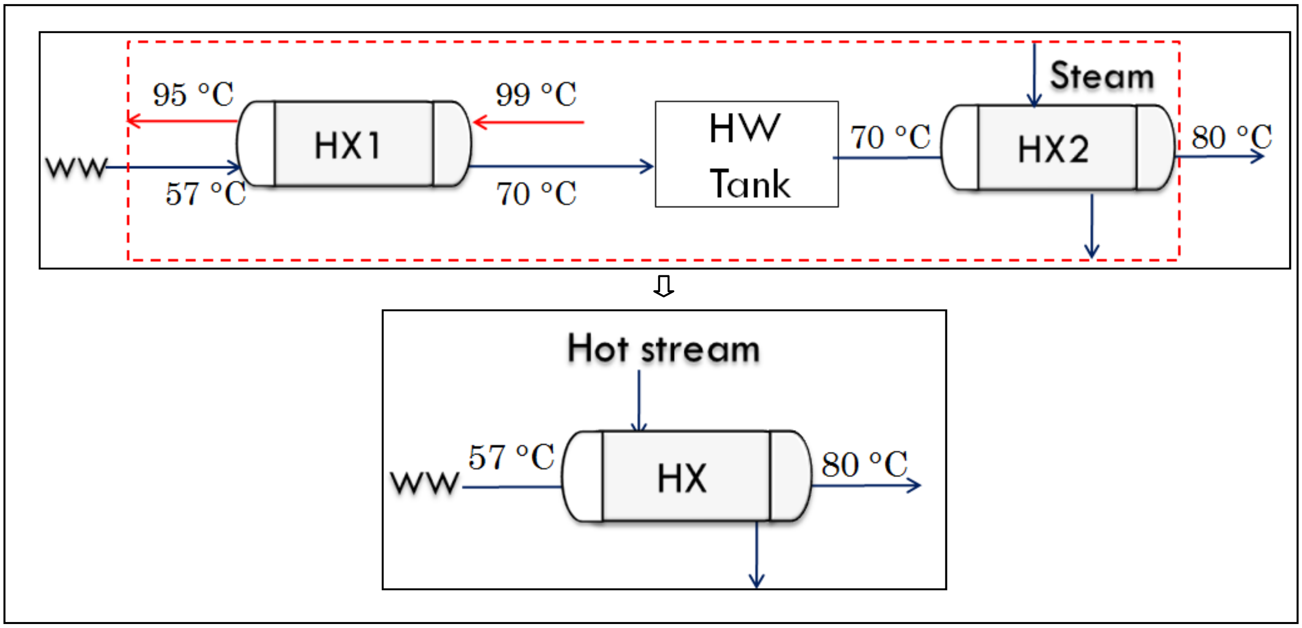


Figure 5-12: Grassroot approach

The data extracted is presented in the following manner:

HX2: $ColdStream = 70\ ℃ \rightarrow 80\ ℃$

Based on the grassroot extraction method, it is possible to achieve savings more than 0.52 GJ/ADT by reutilizing a hot stream that fits the needs of the cold stream in terms of energy load and temperature level and reutilizing the stack gases at a more suitable location. The grassroot approach is applied to the water network which is represented in result section.

5.2.6 Evaluating possible scenarios prior to building composite curves

To build a composite curve there are important points to address. For instance separating water system from the rest of process or integrating it with the overall process on composite curves affects the energy saving potentials. For the case of a Kraft mill with two production lines with interconnection between the two, integration of both line on single composite curves or separating them is worthwhile to evaluate. Moreover, susceptibility of water system to grassroot or retrofit approach is important to investigate. The data extraction step is done to evaluate the viability of each hypothesis and the outcomes are presented in the results section. The hypotheses are described below:

1- Water system - separated or integrated

The first concept revolves around the water system constraints. The question is whether it is more economically beneficial to integrate the water system with the process on one composite curve or to separate them on two different composite curves. If the water system is integrated with the process, the constraints are reduced and potential of saving increases due to higher chances of internal heat recovery. Moreover, separating water from the rest of process results in neglecting the connection of water system with the rest of the process, which might results in lowering the scope of saving and lowering capital costs. A tradeoff between the cost and the savings should imply whether or not the water system should be part of the process composite curve. The comparison takes place between two scenarios; the overall composite curves including water and process is compared with the case of two separate composite curves for process and water individually.

2- Line A and B: separated or integrated

The reference mill undergoing the energy analysis in this study consists of two parallel pulp lines. Each line has its own recovery loop and recovery boilers. An exchange of water and liquor takes place between both lines. Different levels of steam are sent to a common header whereby it is sent to the different departments of the mill. There is the possibility of representing both lines on the same composite curve to treat them as one mill. On the other hand, both lines could be separated and presented in two separate composite curves as two distinct mills. In the latter case, the interconnecting streams represents the exchange of water and liquor between two lines, should be extracted for building the composite curve. From an energy savings point of view, it is better to integrate both lines on the same composite curve. But, if the economic aspect of integrating both lines vs. separating them is not attractive and does not result in a feasible and viable option, then separating the lines is a better and less complicated option.

3- Water system - Grassroot and Retrofit

As explained earlier, constraining the water system with retrofit representation as opposed to non-constraining it with grassroot representation affects the scope of energy savings and the total cost involved. In this study, different grassroot and retrofit representations are available for the water system only. The grassroot approach could not be applied to other parts of the process due to integrity of pulping process. It is critical to use the grassroot approach when the considered equipment has no effect on the integrity of the product. All cooking, washing and drying equipment are necessary and they have direct effect on the process and quality of the produced pulp hence these equipment cannot be replaced or removed. The only section prone to restructuring and being extracted in grassroot approach are the heat exchanger network and water tanks. In order to maximize the potential of saving and minimize capital costs, a tradeoff is done between retrofit and grassroot approach. Figure 5-13 below represents the existing water network for line A and line B.

In the retrofit approach, hot streams and cold streams in each heat exchanger and water tank are extracted. There are 24 points of heat transfer that are extracted and most of the existing heat exchanger are kept intact expect the ones crossing the pinch temperature. In the grassroot approach, the existing design is not taken into consideration and water streams are extracted based on the flow and temperature demand at the point of use. In addition, the process hot streams are also extracted but in the same way as the retrofit approach. However, the existing connections are completely ignored and a new design is proposed. The data for retrofit and grassroot are presented in appendix 2-2.

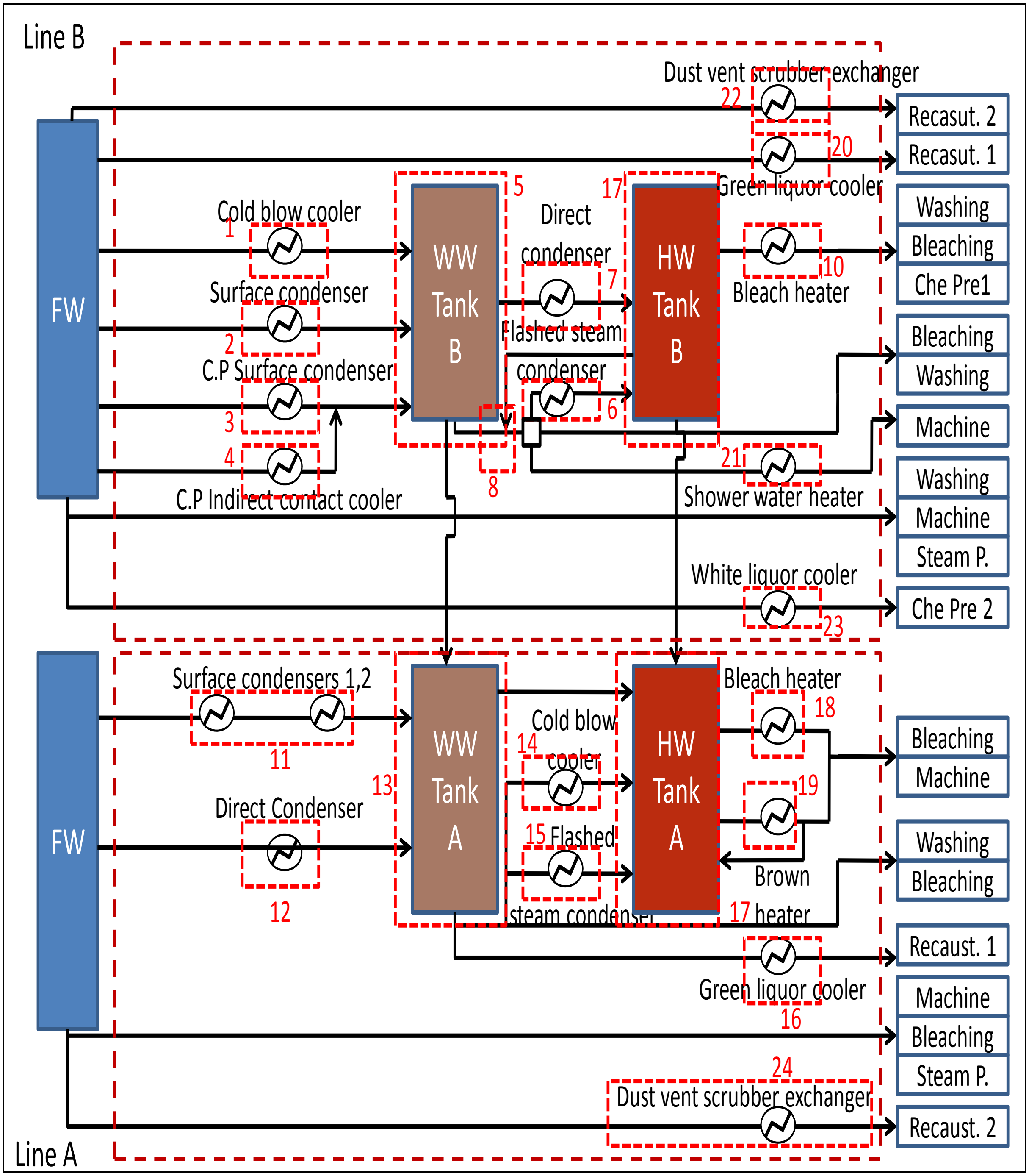


Figure 5-13: Existing water network for line A and line B

5.2.7 Building the composite curves

The set of extracted hot and cold streams are used to build the composite curves on Aspen energy analyzer software. The required inputs into the system include the enthalpy change, temperature in and out, and flow rate. The ΔT_{min} for the process is fixed at 10 °C which is the typical value for the range of 10-15 °C in pulp and paper industry.

The output of the composite curve includes the theoretical minimum heating requirement, theoretical minimum cooling requirement, and the pinch point. In addition, an approximate value of the total heat transfer area is calculated as well as the capital investment, operating cost, and total annual cost. This is evaluated based on a plant life of 5 years and a rate of return of 10%.

The composite curves were built to verify the hypothesis presented in the previous sections as well as to obtain the energy targets for a more efficient process. The new energy target is obtained by reutilization of internal heat and replacing utility steam by other sources of energy. The Initial results of the hypothesis supported by the composite curves are presented in the following section. The list of streams used to build the initial composite curves as well as the actual heating requirement calculation is presented in appendix 2-3.

5.3 Results

5.3.1 Water system– Integrated or separated

As discussed in the methodology, four scenarios are going to be analyzed. Two of the scenarios belong to the case whereby water is on a separate composite curve (Separate) while the other two belong to the case when water is on the same composite curve as the process (Integrated). In addition, the comparison was extended to both line A and line B of the mill.

The criteria used to evaluate the effect of a separate or integrated water system are:

1. MHR: Minimum heating requirement which is obtained by the composite curves
2. MCR: Minimum cooling requirement which is obtained by the composite curves
3. Area: The area of a potential heat exchanger network obtained by aspen energy analyzer.
4. Capital Cost: The capital cost of a potential heat exchanger network resulting in the MHR and MCR

Information and equations used to evaluate the total area and capital cost calculations are presented in appendix 2-4. The composite curves that resulted in the energy, cost and area targets are presented in appendix 2-5.The composite curves results could be summarized in the following bar charts on figure 5-14, 5-15, and 5-16.

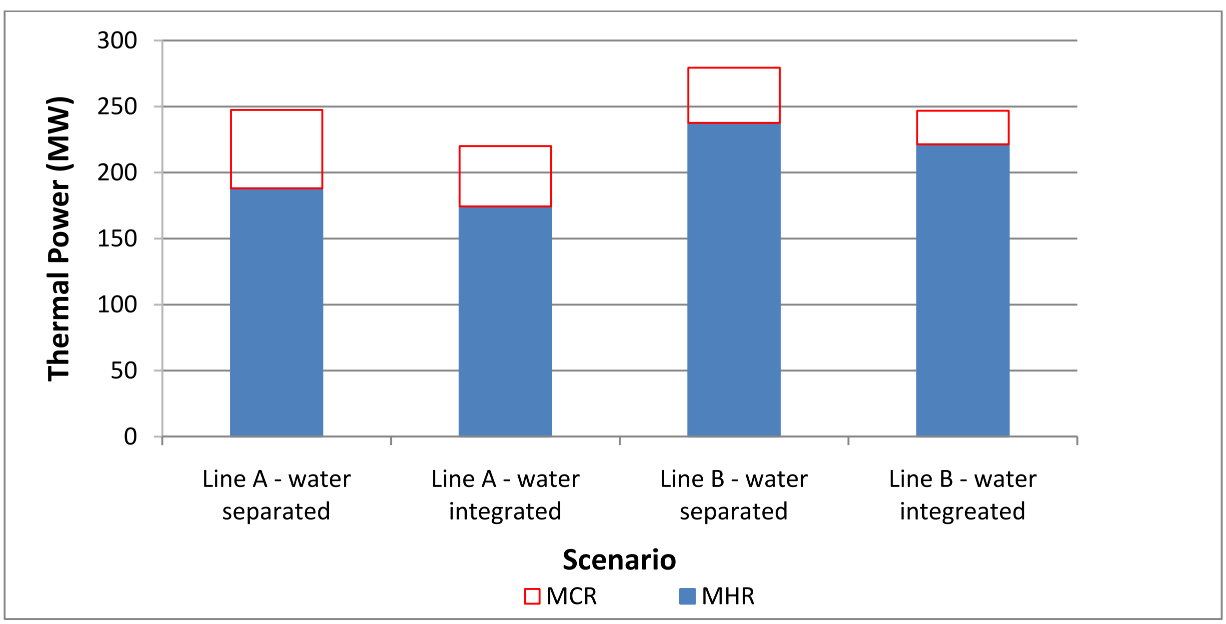


Figure 5-14: Thermal power for Water system - Integrated or separated

Based on the results presented in figure 5-14, by combining the water system and the process on the same composite curve, it is possible to reduce the MHR by 14 MW and 16 MW for line A and line B respectively. This is equivalent to 7% reduction in MHR for each line. To further evaluate the consequences of separating or integrating the water system, the area of potential heat exchanger networks are compared in figure 5-15.

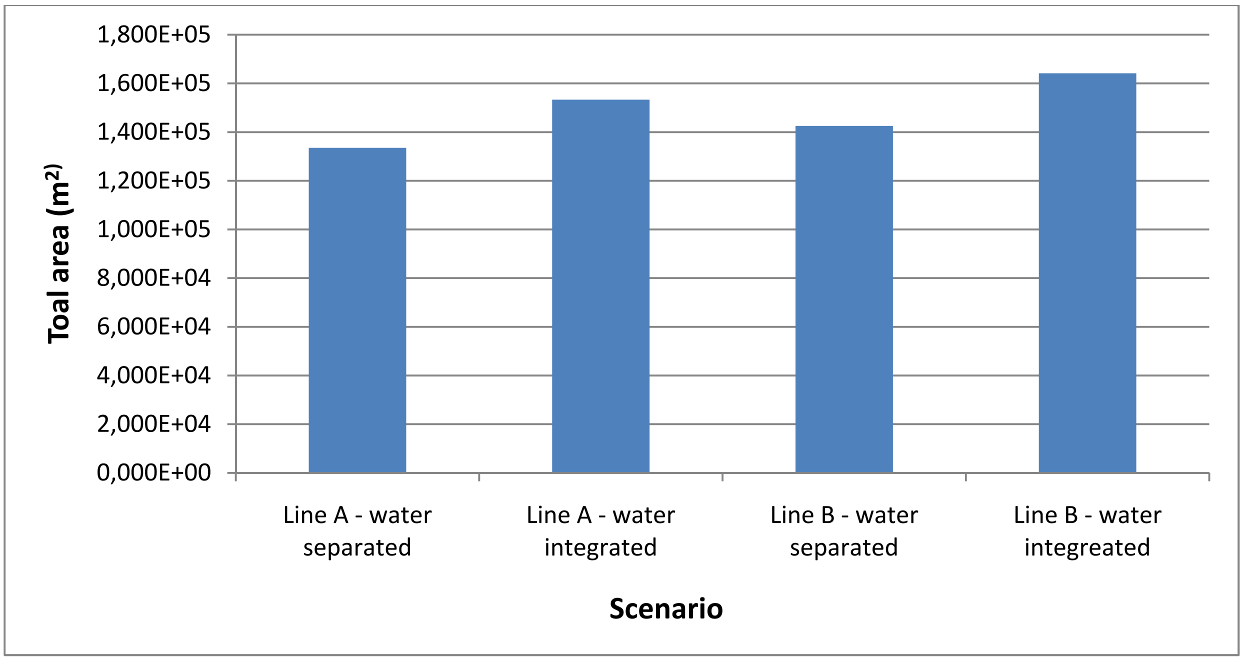


Figure 5-15: Total area for Water system - Integrated or separated

It is evident that the heat exchanger network area increases as the MHR and MCR decreases. This is due to the fact that increasing the internal heat recovery will decrease the MHR but increase the total area needed to perform this heat transfer. The area for both line A and line B increases by 12-13 % for a decrease in MHR and MCR by 7%. At this stage of the analysis, any extra savings is a great opportunity for new potential projects and higher energy reductions. But in order to fully justify the increase in area, a graph of the total capital cost is plotted on figure 5-16.

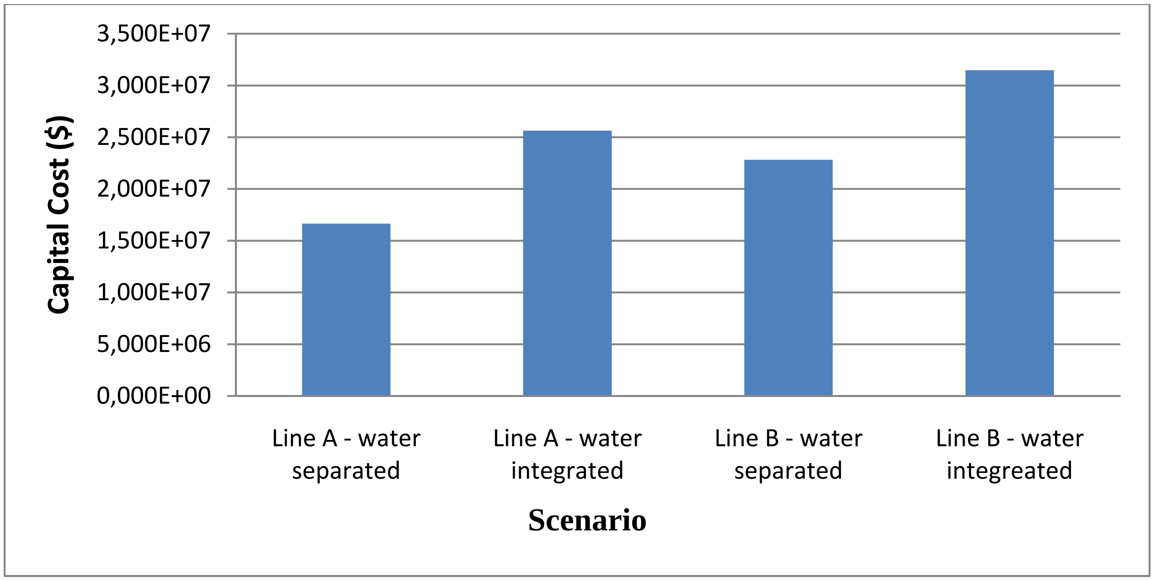


Figure 5-16: Capital cost for Water system - Integrated or separated

Based on the charts, an increase of 25-30 % in total capital cost is noticed when the water system is integrated with the process. At this stage of the analysis an approximate increase of 30 % of capital cost for an increase of 7% in savings is justified and should be considered in the following section of the analysis.

In conclusion, when the water system is integrated with the rest of the process, the MHR and MCR decrease and as a result the area and the capital cost increases. The increase of both capital cost and area are acceptable at this point of the analysis and therefore the integrated water system will be part of the process composite curves in further analysis.

5.3.2 Line A and B – Integrated or Separated

A comparison is made between an integrated system that is formed when both line A and line B is on the same composite curve. Theoretically, this will increase the capacity of internal heat recovery due to the increase of potential connections between both lines. A separate system whereby Line A and Line B are plotted on two composite curves will be compared to the integrated system.

The composite curves of these systems are plotted on Aspen energy analyzer® to produce the energy targets, total areas, and capital cost approximations. These composite curves are presented in Appendix 2-6. In order to summarize the results of the composite curves, the following charts are plotted and presented on figure 5-17.

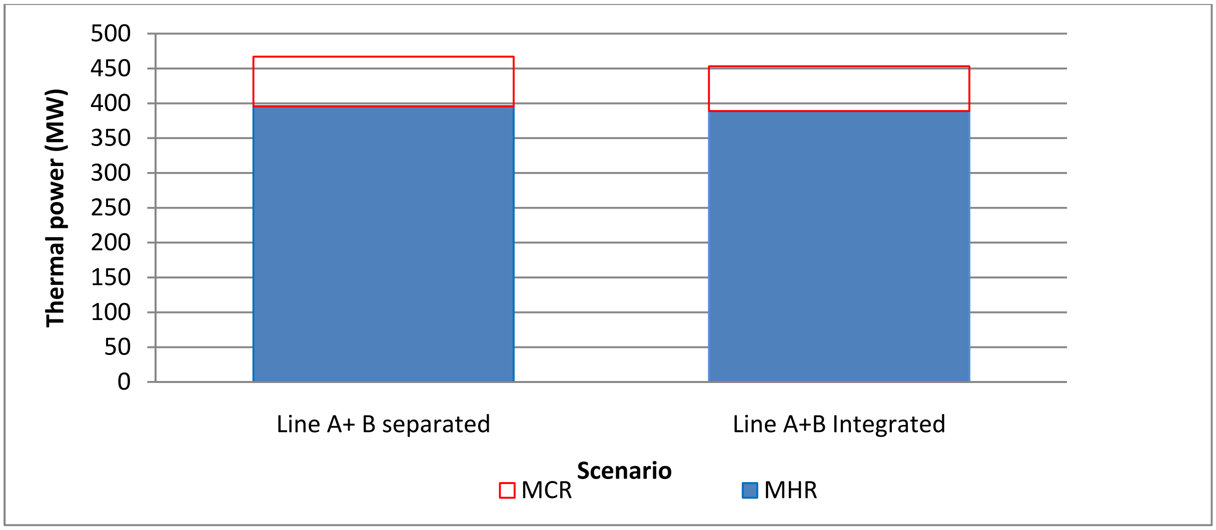


Figure 5-17: Thermal power for Line A and B – Integrated or Separated

It is evident that by integrating both lines on the same composite curves, MHR and MCR decrease by 2%. This increase in total internal heat recovery is very small when compared to the case where water system is separate or integrated with the process. Nonetheless, this could potentially increase the total internal heat recovery. Further analysis of the scenario is required by examining the total area and capital cost.

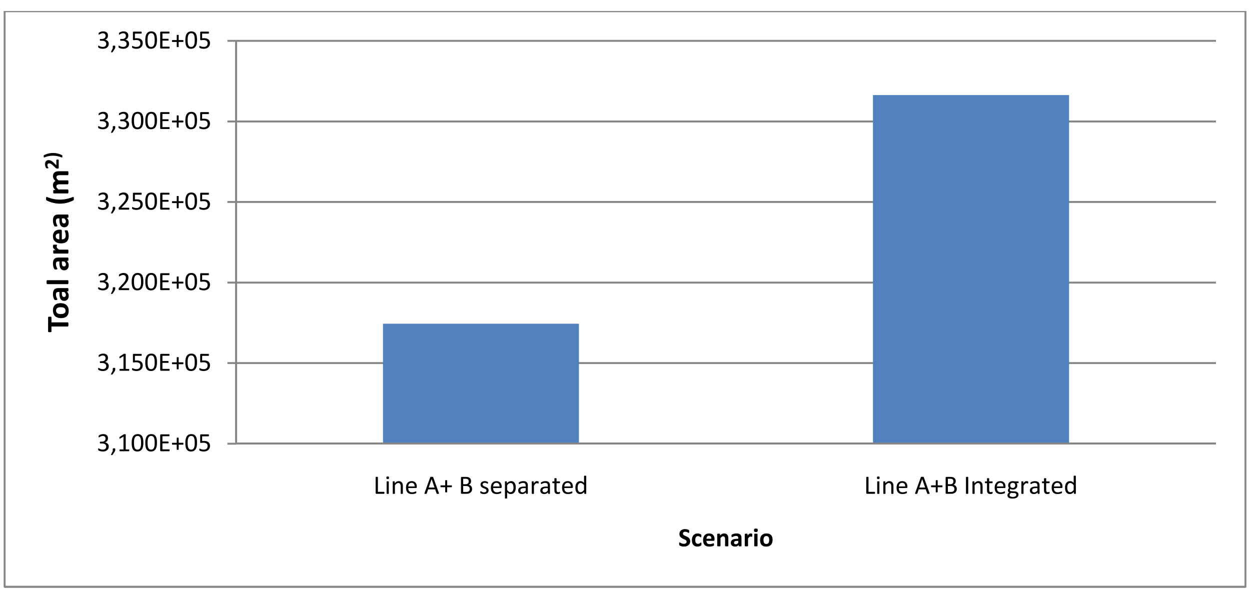


Figure 5-18: Total area for Line A and B – Integrated or Separated

Figure 5-18 represents the total area for an integrated system vs. separated system. By combining both lines the area increases by 5 % for an increase in total internal heat recovery of 2%. Based on these results, it seems like that there is a potential benefit of including both lines on the same composite curve.

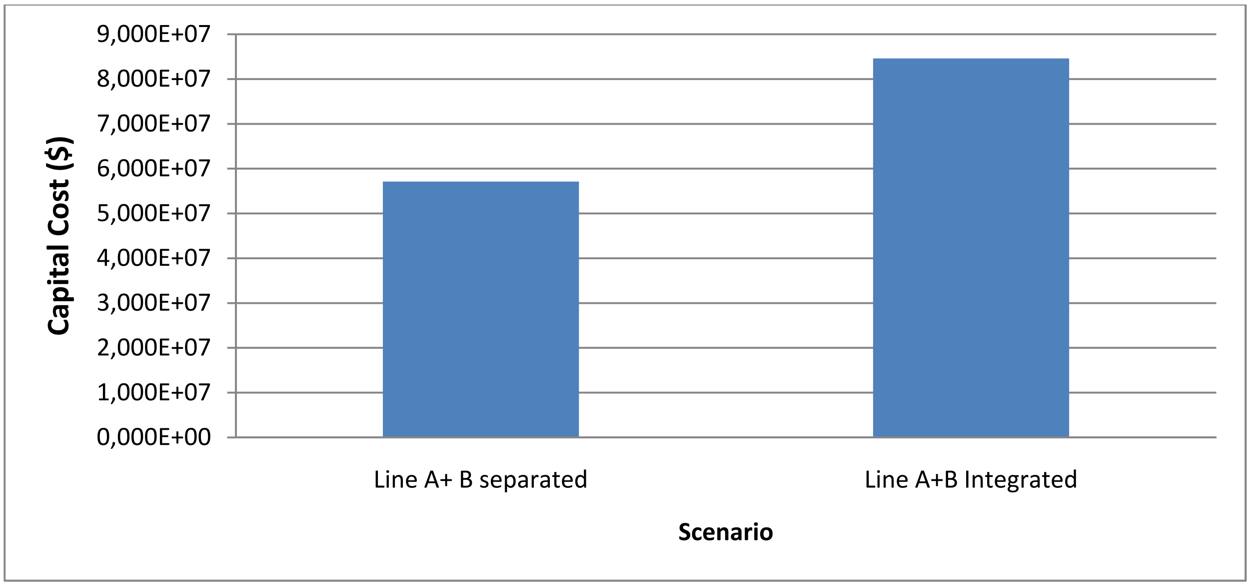


Figure 5-19: Capital cost for Line A and B – Integrated or Separated

Based figure 5-19, it is noticed that there is a significant increase in capital cost due to the energy saving of 2%. It shows that the capital cost increases by 50% due to the integration of both lines on the same composite. Hence, integrating both lines on the same composite curve not only increases the complexity of the system, it doubles the capital cost involved with the design thus making this option less viable. As a result both lines are analyzed independently.

5.3.3 Grassroot approach vs. Retrofit approach

In order to explain the role of a grassroot and retrofit data extraction on the process design and energy targets, four composite curves were plotted. Figure 5-20 represents the method of obtaining the four curves:

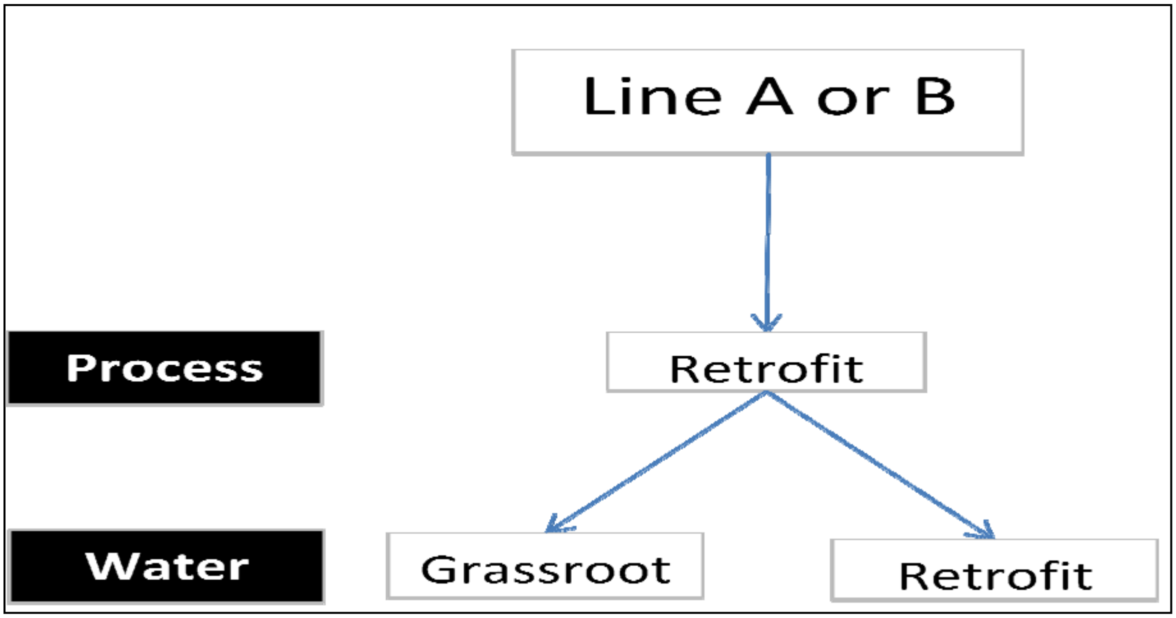


Figure 5-20: Method of obtaining grassroot and retrofit schematic

Two composite curves were developed for each line. The process streams were always extracted in retrofit approach while the water network streams were extracted either in retrofit or grassroot approach. The focus of this analysis is to examine two extreme cases whereby the existing design could be modified "retrofit" or where by a completely new design is proposed "grassroot". Therefore it could be said that the retrofit approach is a constrained approach which will have a lower scope of energy savings when compared to the grassroot approach which is a non constrained approach. The criteria for assessing the impact of these two different approaches would be:

1- Minimum Heating Requirement
2- Minimum Cooling Requirement

In addition, the total area, capital cost, operating cost savings, and payback period are important criteria that will be evaluated in chapter 6.

The four composite curves were plotted on Aspen Energy Analyzer® and are presented below. First of all, line B thermal composite curves and grand composite curves are going to be discussed and analyzed. Line A analysis will be presented after the analysis for line B.

5.3.3.1 Thermal Composite curves and grand composite curves – Line B

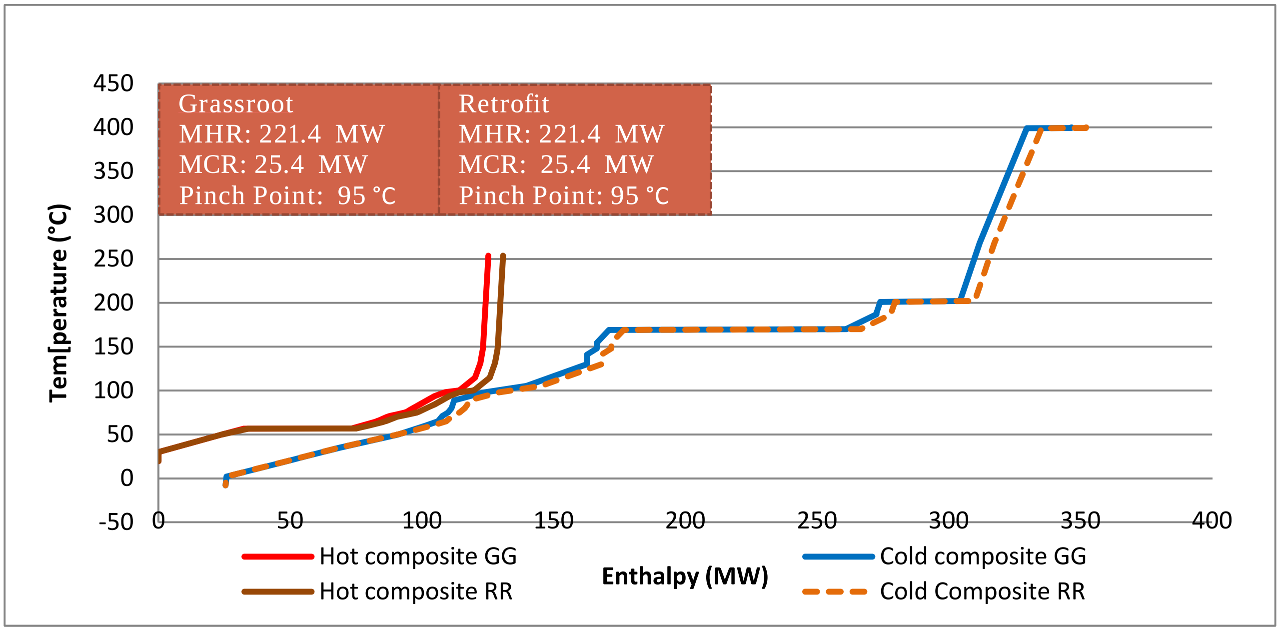


Figure 5-21: Line B - Grassroot vs. Retrofit composite curves

Based on the composite curves for grassroot and retrofit presented in figure 5-21, Minimum Heating Requirement (MHR) is 221.4 MW while the Minimum Cooling Requirement (MCR) is 25.4 MW. Both representations results in the same minimum heating requirements and cooling requirements. This indicates that in this specific line a constrained or non constrained approach in the water system does not affect the final energy target. This is due to the fact that the hot streams and cold streams below the pinch did not change and therefore the scope of internal heat recovery did not change. It is also noticed that above the pinch, there is a shift in both the hot streams and cold streams in the retrofit case. This phenomenon occurs due to the addition of streams mixing in a non isothermal manner in hot water and warm water tanks to the retrofit composite curve streams. In the grassroot approach, these streams were not included in the composite curves.

In order to have a better understanding of the nature of these curves, the grand composite curves are plotted and presented in figure 5-22:

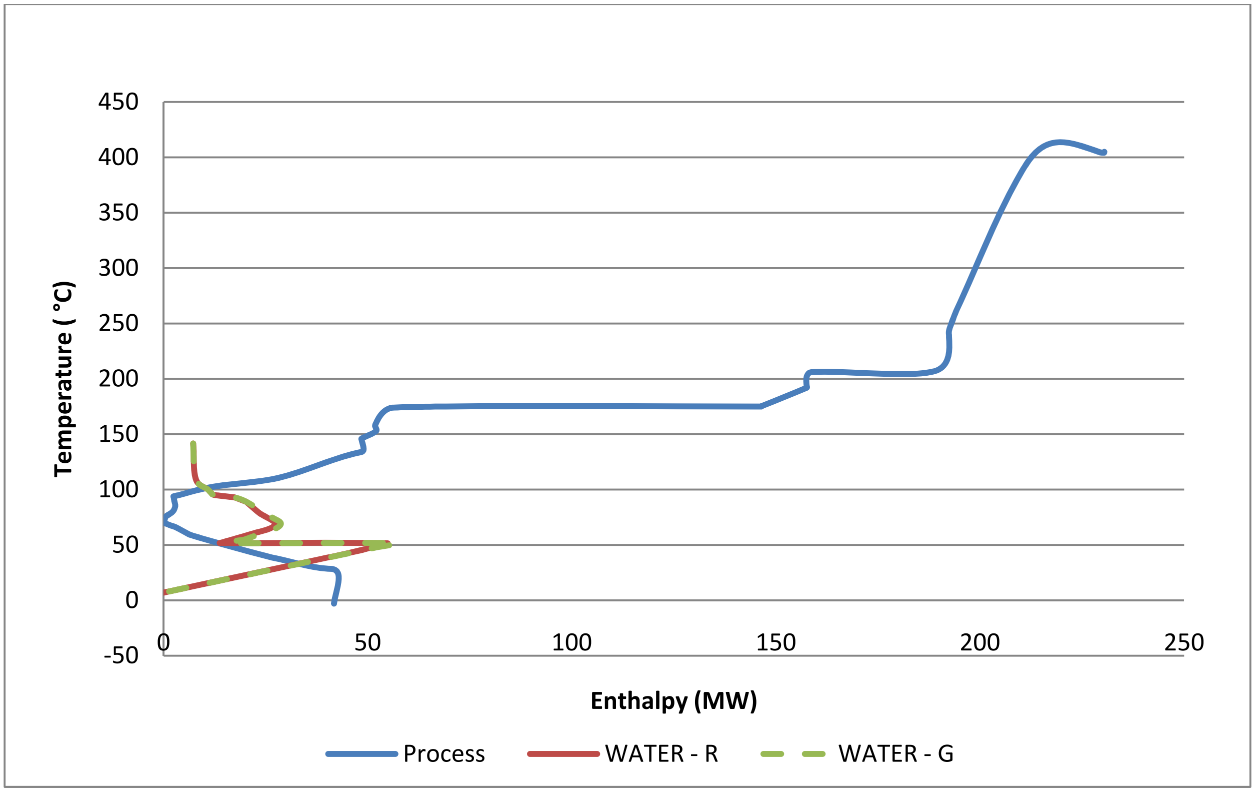


Figure 5-22: Grand composite curve - Line B Process vs. water

The blue curve represents the grand composite curve for the process which includes all the process streams except for the cold water system. The green dotted curve is the water system streams extracted in a grassroot manner. The red curve is the water system streams extracted in a retrofit manner. It seems like both retrofit and grassroot curves are identical but a closer look is required to identify if there are any discrepancies between both curves under the pocket of energy occurring at the pinch temperature. An interesting fact to note is that whether the water system is presented in a grassroot or retrofit, the total energy load or cold demand remains the same but the interaction between hot process streams and cold water streams changes. This change will result in a different heating requirement and cooling requirement and therefore a different design. Figure 5-23 provide a closer look on the pocket for the process, grassroot and retrofit grand composite curves:

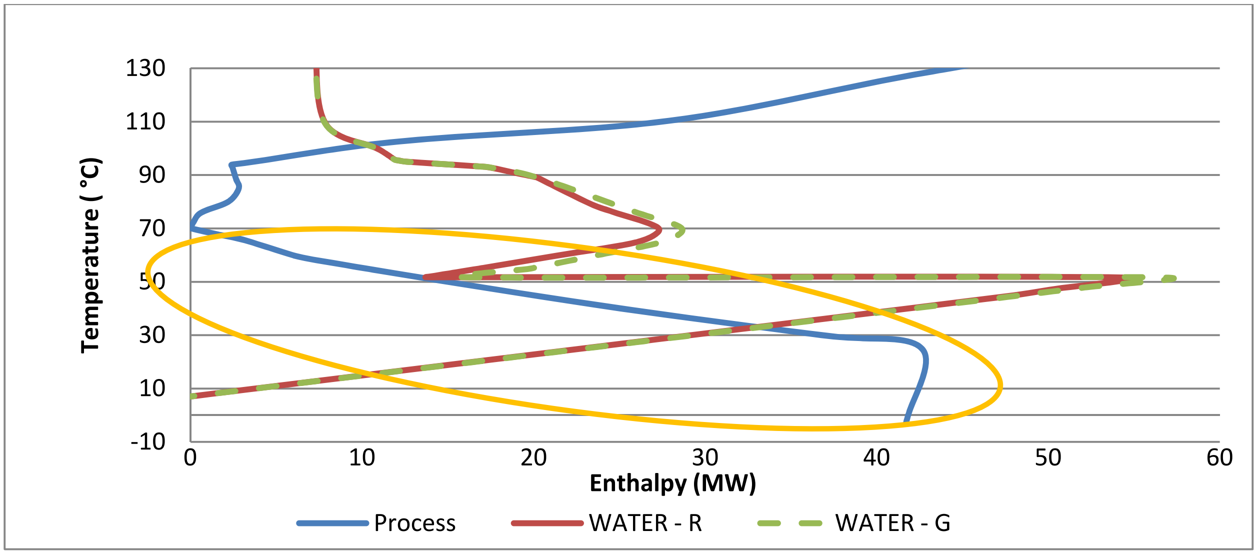


Figure 5-23: Grand composite curve - Close up Line B

By focusing on the highlighted zone which indicates the heating capacity (required cooling of process streams) below the pinch, we notice that both grassroot streams and retrofit streams are identical. Therefore the interaction between the hot streams of the process and the cold water streams in both representations is the same. The difference in grassroot and retrofit curves starts to appear after 50 °C where the shape and the heating levels of the curves changes slightly. Since the change occurs above the level of the heating pocket available in the hot streams, it doesn't have an effect on the overall heating and cooling targets. As a result the grassroot for line B was not carried out and the focus was mainly on the retrofit approach.

The results for line A were significantly different and the influence of grassroot approach was more apparent. The thermal and grand composite curves are presented in figure 5-24:

5.3.3.2 Thermal Composite curves and grand composite curves – Line A

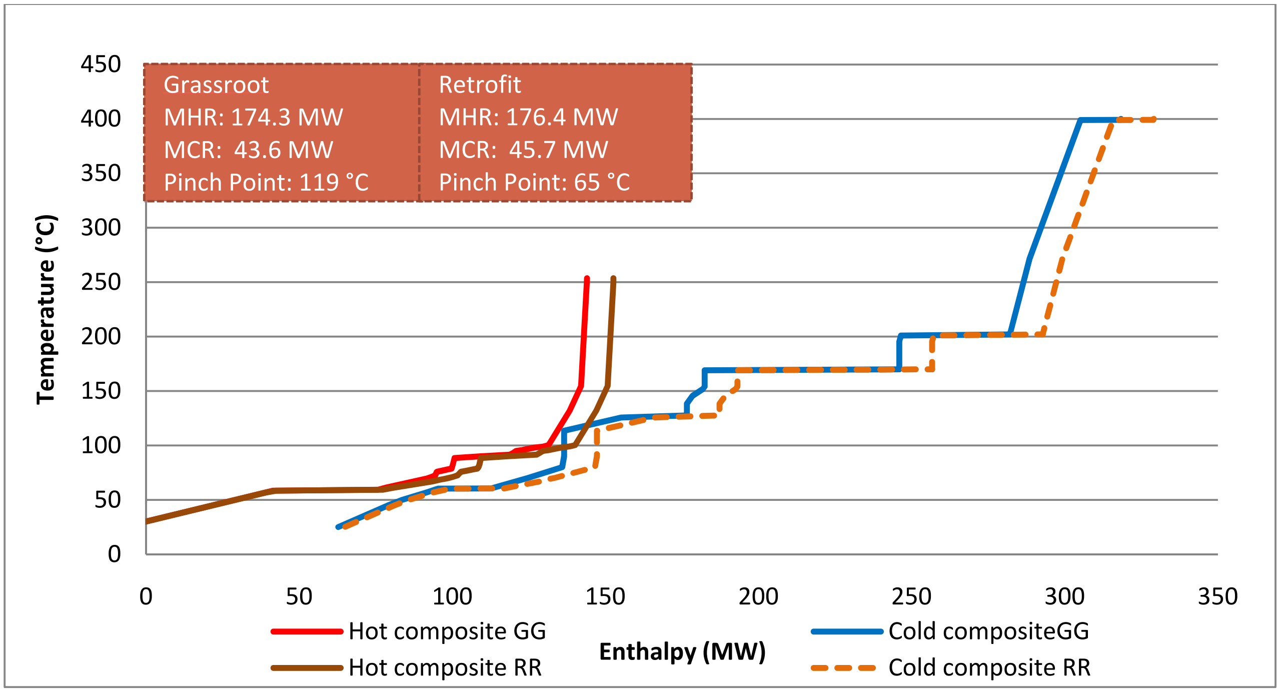


Figure 5-24: Line A - Grassroot vs. Retrofit composite curves

There is a difference of 2 MW in MHR and MCR between grassroot and retrofit curves. In addition, the pinch temperature changes significantly from 119 °C to 65 °C between the two representations. This change certifies the initial assumption whereby the grassroot approach will result in a lower minimum heating requirement due to increasing the possibilities of internal heat recovery. The same phenomenon that occurred in line B can be seen in the thermal composite curves as well. The retrofit representation of the water system leads to the shifting of hot and cold composite curves to a higher enthalpy. This is due to the addition of more streams under the form of non isothermal mixing streams in warm water and hot water tank. The difference between line B phenomena is that the shift in both curves did not result in the same MHR and MCR as the grassroot case. In order to understand the essence of this reduction in MHR and MCR, the grand composite curves of the process and water system should be examined and are presented in figure 5-25.

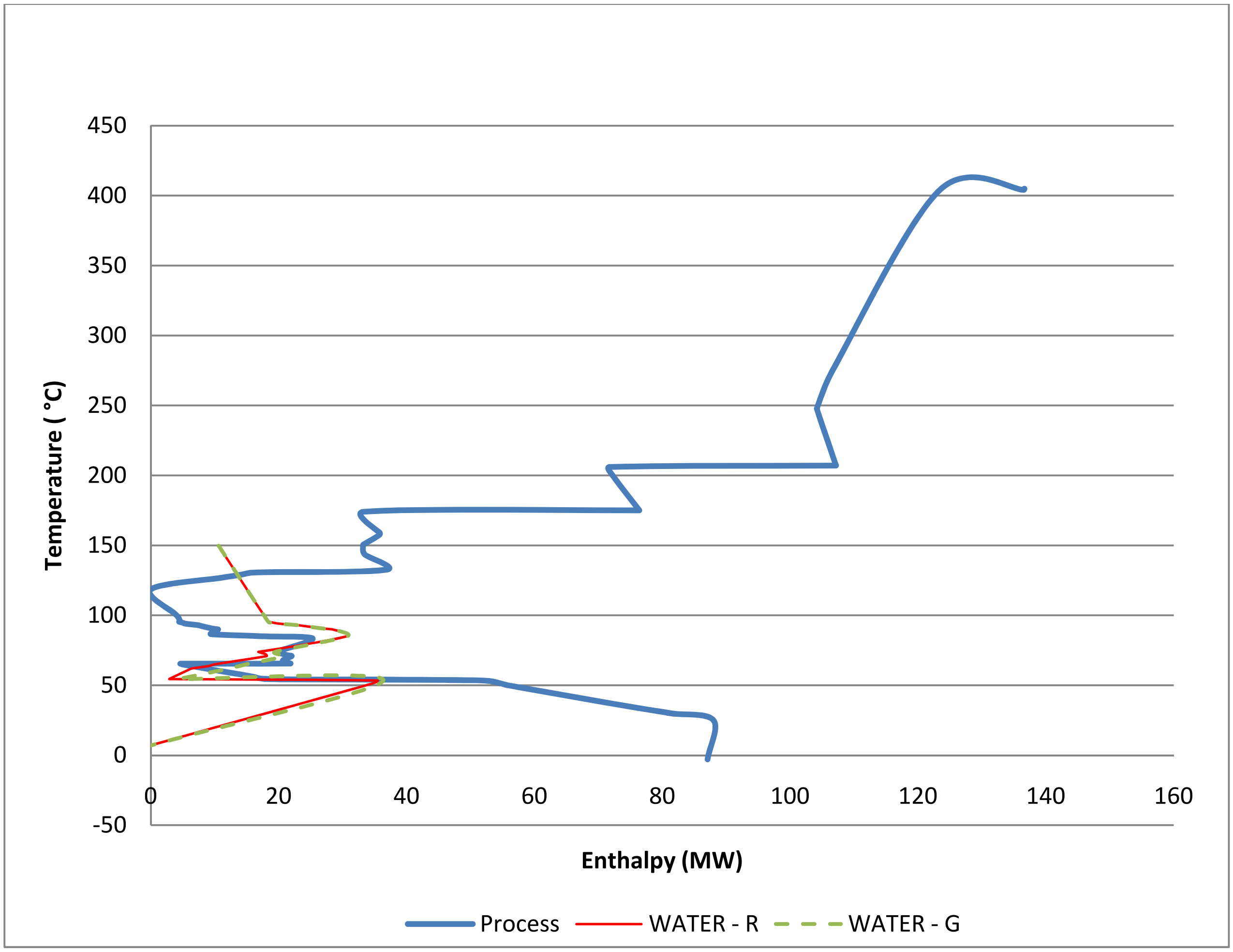


Figure 5-25: Grand Composite Curve - Process vs. water Line A

The blue curve represents the grand composite curve for the process which includes all the process streams as well as the hot process streams used to heat the water system. The green dotted curve is the water system streams extracted in a grassroot manner. The red curve is the water system streams extracted in a retrofit manner. From this view, it seems like both retrofit and grassroot curves are different under the second hot pocket. A closer look is required to identify if the water grand composite curve have a different shape under the hot pocket. The following curves in figure 5-26 provide a closer look on the pocket for the process, grassroot and retrofit grand composite curves:.

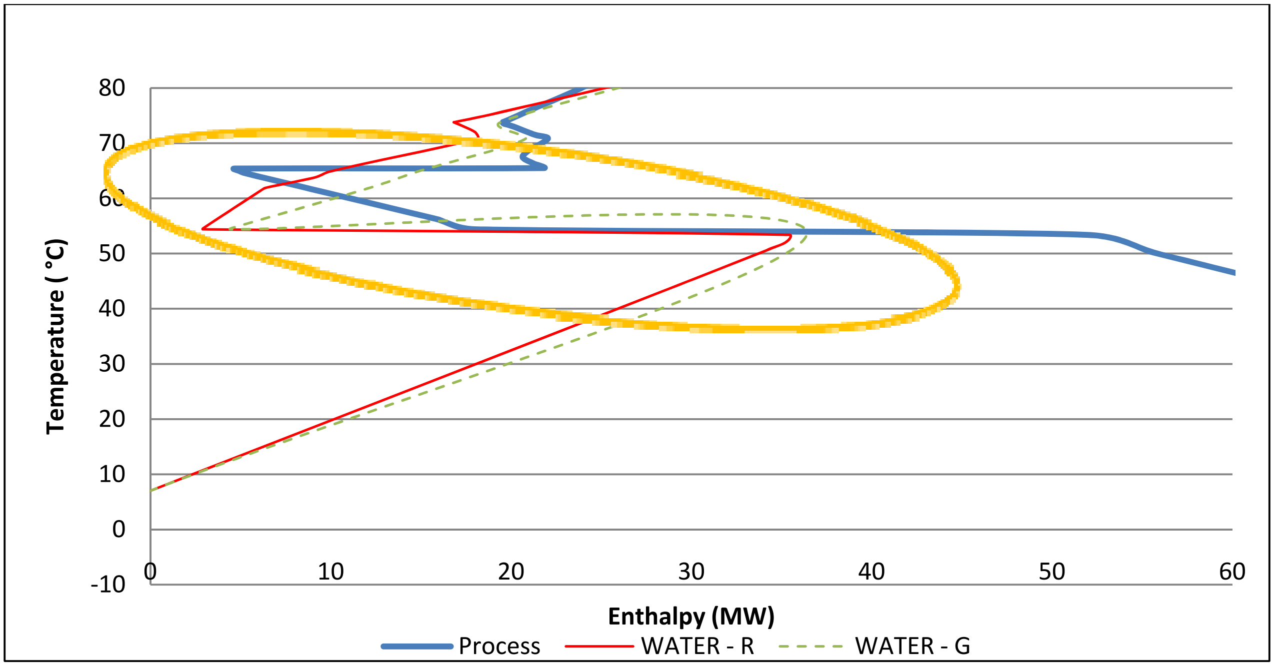


Figure 5-26: Grand composite curve - Close up Line A

Unlike what was noticed in line B, the retrofit and grass root water curves are not identical under the hot pocket. The hot pocket mainly consists of hot process streams that need to be cooled. The best scenario is to use these streams to heat cold water streams and reduce the dependence on steam as a source of energy. The grassroot composite curves have a lower slope that intercepts the process grand composite curve at a further point under the heat pocket. In addition at around 55 °C, the retrofit composite curve extends by 2 MW to the left side of the chart. This means that the required 2 MW cannot be supplied by the process hot stream and need utility steam as an energy source. On the other hand, the grassroot cold composite curve remains under the hot pocket streams and therefore requires no utility steam. This discrepancy explains the reason why there is a difference of 2 MW between the MHR and MCR for retrofit and grassroot composite curves.

Based on the discussion presented above, grassroot approach will result in superior heat exchanger design with higher energy savings. This new design will require a larger surface area and therefore a larger capital cost than a retrofit design. In chapter 6, a heat exchanger network will be designed for the grassroot case and the retrofit case to determine whether it is economically viable to go through a complete restructuring of the mill water network to save an extra 2 MW of energy. A complete comparison will be based on the total area, capital cost, operating cost savings and payback period. For line B a retrofit approach will only be proposed due to the lack of savings between both grassroot and retrofit representations.

Table 5-3 includes a summary of the composite curves results and the energy savings involved.

Table 5-3: summary of the composite curves results

	Line A		**Line B**	
Data	**Grassroot**	**Retrofit**	**Grassroot**	**Retrofit**
MHR (MW)	174.4	176.4	221.4	221.4
MCR (MW)	43.6	45.7	25.5	25.4
Pinch temperature (°C)	118.5	65	95	95
AHR (MW)	196		249	
Savings (MW)	21.6	19.6	27.2	27.3
Savings (%)	11	10	11	11

5.3.4 Identifying all potential energy saving projects

At this stage new projects should be identified in order to reduce the minimum heating requirement thus increasing the energy savings. In the initial composite curves, points of heat transfer that uses non constrained steam were identified and included in the curves. Extra energy saving projects should be identified by focusing on the elimination or reduction of directly injected steam or non isothermal mixing points (type 4 constraints). Based on the constraint analysis schematic, type 1c direct steam injection constraints will shift to type 4 non isothermal mixing elimination constraints. This will result in huge energy savings in different areas of the mill.

In order to detect the exact location of these projects, a good knowledge and understanding of the process is required. This was discussed at the beginning of the constraint analysis strategy section. None the less, pulp line screening tool as well as tank screening tool helped in identifying areas of dilution or steam injection as well as non isothermal mixing in tanks. The idea is to thoroughly examine each non isothermal mixing point and to find an alternative by using internal heat recovery to reduce the destruction of high quality energy. An important thing to note is that at this stage only the cold demand or the stream that needs to be heated and the energy savings are known. The identification of the heat source in each project will take place after building the final composite curve and heat exchanger network. More information regarding the configuration of the projects will be presented in chapter 6. The projects that were identified for line A and line B are presented in table 5-5 and 5-6 respectively.

Table 5-4: Line A potential energy saving projects

Project number	Project name	Steam Savings (MW)
1	Bleach heater	7.37
2	Brown heater	3.24
3	Boiler Air Heater	3.16
Non isothermal mixing elimination projects		
4	Deaerator (make up water)	6
5	Injection 1 (Washer 15)	1.7
6	Injection 2 (Washer 35)	2.7
7	Injection 3 (Washer 45)	2.94
8	Injection 4 (Washer 55)	1.6
Total		**28.7 (15%)**

Table 5-5: Line B potential energy saving projects

Project number	Project name	Steam Savings (MW)
1	Boiler Air Heater	7.59
5	Bleach heater	2.82
	Direct Condenser	4.4
Non isothermal mixing elimination projects		
2	Deaerator (make up water)	14.25
3	NIM in Washing Dilution conveyer	0.81
4	NIM in White Water tank	2.93
6	Injection 1 (washer 15)	3.47
7	Injection 2 (washer 35)	2.17
8	Injection 3 (washer 45)	1.88
9	Injection 4 (washer 55)	2.54
Total		**42.87 (16%)**

5.3.5 Building the refined composite curves

The last step is to build the refined composite curves by applying the following changes:

1. Addition of Non isothermal mixing elimination projects

Non isothermal elimination will replace or reduce the load from direct injection constrained steam. A shift will occur in the position of type 1c. high temperature levels cold streams. These streams will be replaced by a type 4 low temperature levels cold streams. As a result the minimum heating requirements is reduced. One will expect that the reduction in minimum heating requirement will be equal to the energy savings from the project. This is the case when enough heat is available in hot streams to satisfy the need of the new added cold streams. In the case whereby no reduction in minimum heating requirement is noticed, it is possible to say that the energy available is inadequate to satisfy the needs of the new projects. As a result the proposed potential projects would not be viable. The list of streams for the non isothermal mixing projects is in appendix 2-7.

2. Refining the effluent streams and gases

One more look is required to assess the energy load and temperature levels of the effluents.

Effluents are produced in different sections in a Kraft mill. Depending on the location in the mill, these effluents could be in liquid form or gas form. Also, the temperature level and energy content of these effluents is dependent on the equipment or sections where they are produced. In practice, these effluents are treated and released into the environment. In some cases, these effluents are treated and reused in the process to recapture some of their energy before being sent to the environment. Therefore it is important to assess the quality and quantity of energy in these effluents in order to use them as a new heat source in the process.

In line A the total number of hot streams that fit the initial criteria mentioned in the guidelines for constraint analysis section are 27. Out of the 27 hot streams, there are 14 distinct effluent streams in the process with high potential of energy reutilization. In line B, the total number of hot streams is 32. 16 of these 32 hot streams are effluents with potential for heat recovery. The idea is to screen the effluents and determine which ones could be practically used in a heat exchanger. Some effluents that had an energy load small than 0.4 MW were removed from the list. In addition some effluents that were impossible to be used due to technical difficulties were removed as well. The list of effluents after refinement is presented in appendix 2-8.

3. Addition of stack and flue gases

The last step before finalizing the curves is to include the recovery and power boilers stack gases. This is an important step to assess the overall ability to save energy in the process.

The curves before and after refinement for line A and line B in retrofit and grassroot are presented below and compared in terms of minimum heating requirement, minimum cooling requirement, and energy savings.

5.3.6 Line A refined composite curves

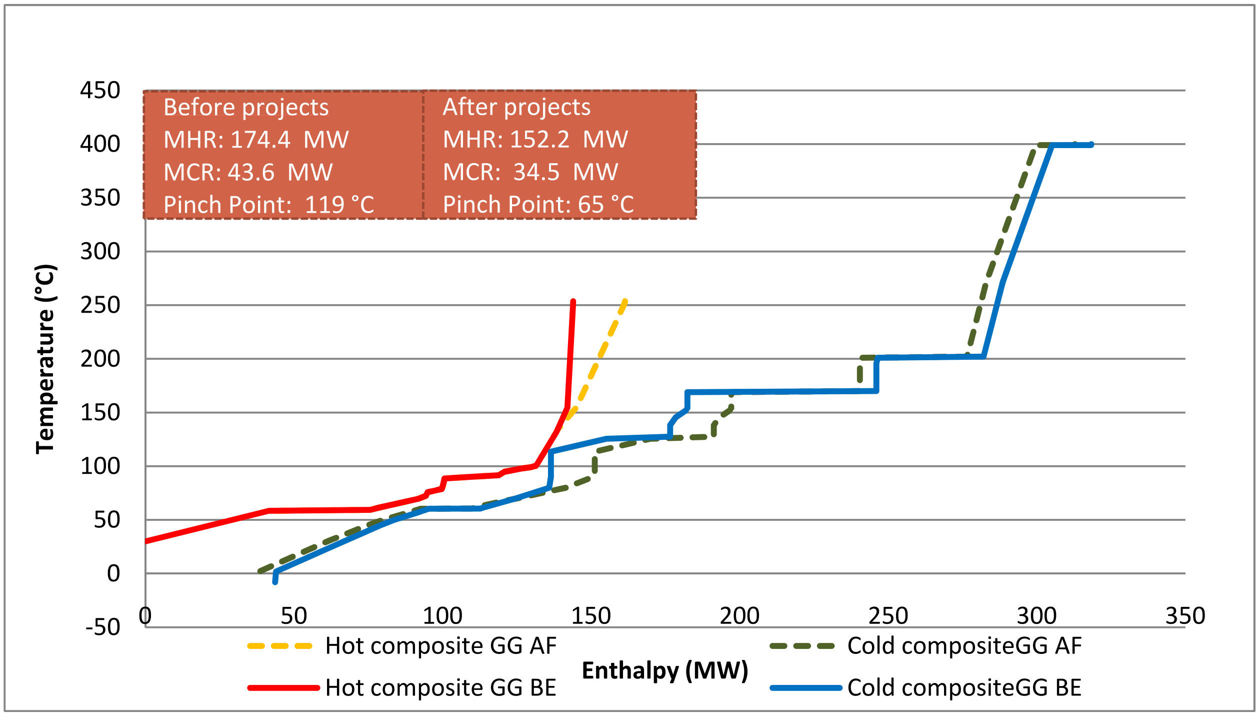


Figure 5-27: Line A - Grassroot After and before refinement

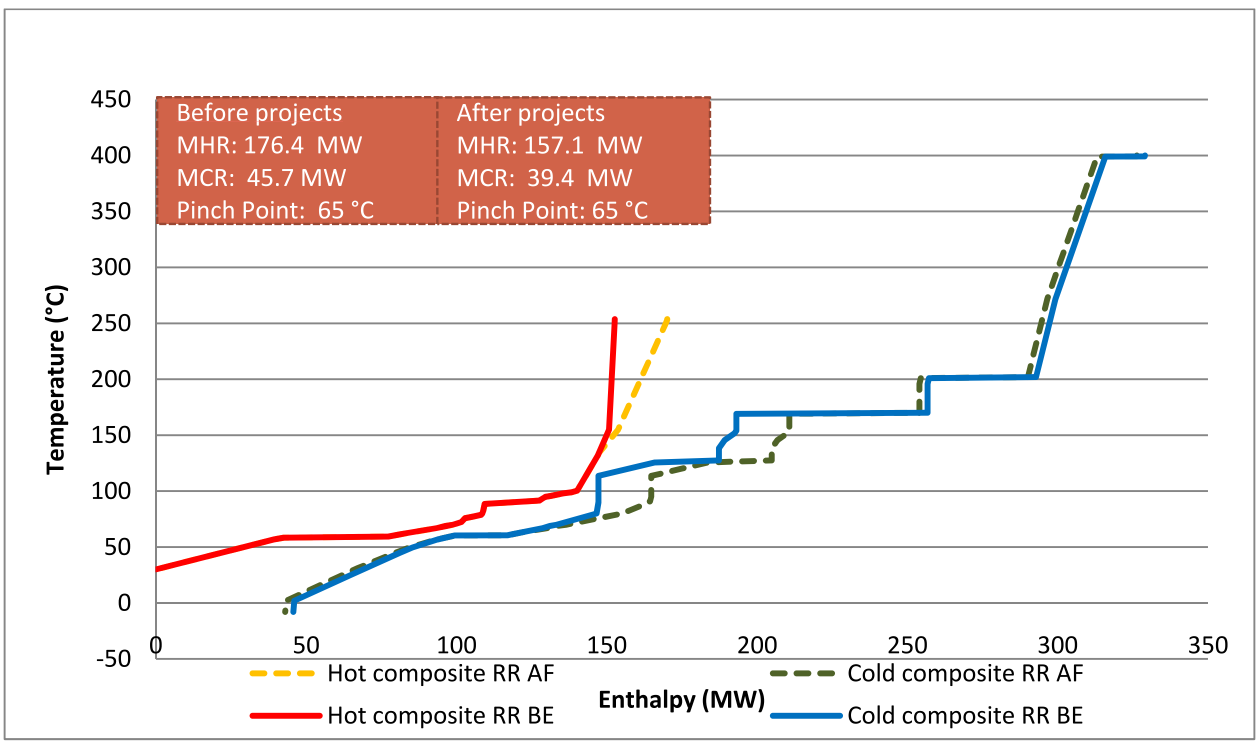


Figure 5-28: Line A - Retrofit After and before refinement

5.3.7 Line B refined composite curves

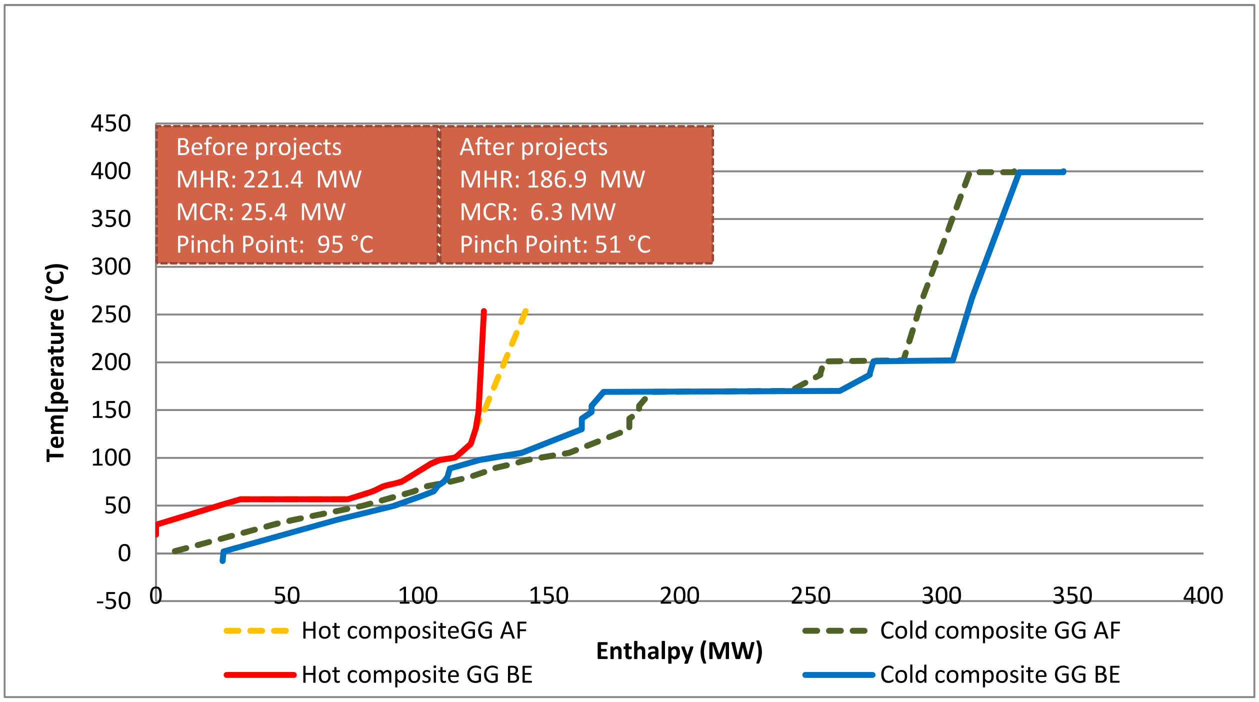


Figure 5-29: Line B - Grassroot After and before refinement

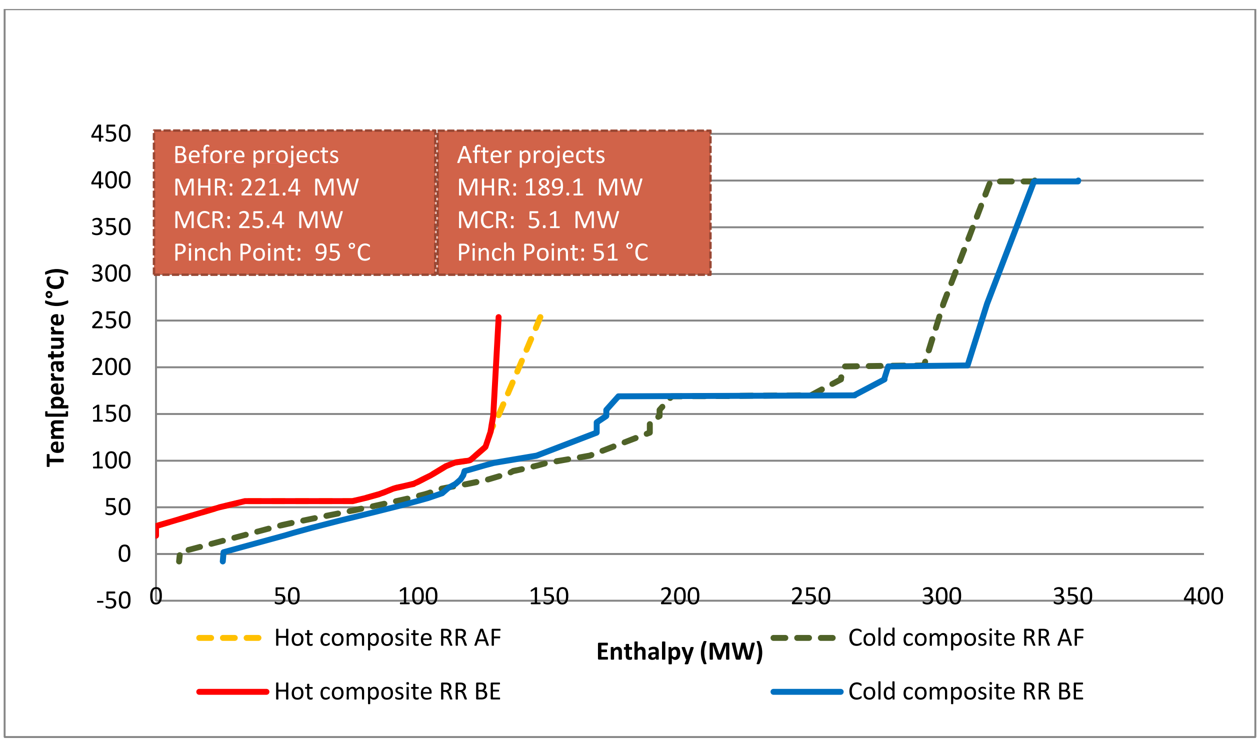


Figure 5-30: Line B - Retrofit After and before refinement

5.3.8 Summary of refined composite curves

Based on figures 5-27 and 5-28, it is evident that in Line A an increase of 10 -11% in energy savings was seen in retrofit and grassroot after the refinement of the curves. It is also clear that there is a difference of 2% in savings which is equivalent to 5 MW between both approaches.

Table 5-6: Summary of refined composite curves results - Line A

Line A	**Grassroot**		**Retrofit**	
Data	**Before Projects**	**After Projects**	**Before Projects**	**After Projects**
MHR (MW)	174.4	152.2	176.4	157.1
MCR (MW)	43.6	34.5	45.7	39.4
Pinch temperature (°C)	118.5	65	65	65
AHR (MW)	196		196	
Savings (MW)	21.6	43.8	19.6	38.9
Savings (%)	11	22	10	20

In line B figure 5-29 and 5-30, huge savings between the refined curves and the initial composite curves is noticed. Based on the refined composite curves, 24-25 % of theoretical energy savings is achieved. What was noticed is that the difference in savings between grassroot and retrofit after refinement remained relatively small (2 MW). As a result a heat exchanger network in grassroot approach is not going to be proposed in chapter 6.

Table 5-7: Summary of refined composite curves results - Line B

Line B	**Grassroot**		**Retrofit**	
Data	**Before Projects**	**After Projects**	**Before Projects**	**After Projects**
MHR (MW)	221.4	186.9	221.4	189.1
MCR (MW)	25.5	6.3	25.4	8.5
Pinch temperature (°C)	95	51.0	95	51.0
AHR (MW)	249		249	
Savings (MW)	27.2	61.8	27.3	59.6
Savings (%)	11	25	11	24

In conclusion, it is safe to say that a grassroot approach in Line A will result in higher theoretical energy savings. A decision on whether it is more economically beneficial to design a heat exchanger in grassroot cannot be made at this point in the analysis. For line A, three heat exchanger networks in retrofit and grassroot are going to be designed and compared from an economic point of view to make a final decision regarding both representations.

CHAPTRE 6 HEAT EXCHANGER NETWORKS AND ENERGY SAVING PROJECTS

6.1 Introduction:

In this chapter, the idea of proposing energy saving projects and developing a new heat exchanger network will be discussed. More information regarding the energy saving projects that was presented in the guidelines chapter will be addressed. The steps required to build a heat exchanger network with the new projects will be presented. Generally, four major steps in building a heat exchanger network are identified and presented in the following order.

1- Building existing heat exchanger network for both lines

2- Evaluating the energy violations in the existing networks

3- Building new heat exchanger networks by correcting the energy violations and implementing energy saving projects into existing heat exchanger network

4- Economic analysis based on capital cost, operating cost savings, and a simple payback period

In line A, three heat exchanger networks at three constraint levels are built and analyzed. The constraint levels are “Retrofit – low savings”, “Retrofit – medium savings”, and “Grassroot – high savings”. The flexibility of the design and the number of targeted/modified heat exchangers increase as the constraints are reduced in each level. Reducing the constraints will result in higher savings and therefore increase the total area and capital cost of the proposed heat exchanger network. A tradeoff between constraint levels, energy savings, and capital cost is a key issue to be looked at while choosing the best heat exchanger network design. In line B, “Retrofit- medium savings” heat exchanger is built and analyzed. The four major steps for building and analyzing a potential heat exchanger network are discussed in details below.

6.2 Building the existing heat exchanger network

The first step of building a new heat exchanger network is to construct and evaluate the existing heat exchanger network. The stream information required for this task has already been extracted while building the composite curve. The stream information needed includes stream temperatures, specific heat capacities and duties of all the streams in the heat exchangers.

The existing heat exchanger network is built on two different interfaces which are Aspen energy analyzer and Microsoft Excel. The heat exchanger network is constructed using the composite curves built on Aspen energy analyzer®. This will enable the identification and the quantification of any cross pinch violation in the heat exchangers. In addition, a qualitative analysis is applied on Microsoft Excel to identify heat exchangers with "Criss-cross" heat transfer. This will be explained in more depth in step 2. The configuration and the information of the heat exchanger networks for line A and line B are presented below.

6.2.1 Existing heat exchanger network – Line A

The heat exchanger network for line A is separated into process network and water network. The process network includes all the heat exchangers that have no interaction with the heating of water streams. These heat exchangers are:

- Boiler air heater
- Black liquor heater
- Lower and upper cooking liquor heaters
- Glycol loop heaters

The water network includes all the heat exchangers responsible for heating water streams before it is consumed in different units in the process. Figure 6-1 includes the heat exchangers in the process network and water network.

Figure 6-1: Existing heat exchanger network - Line A

A list of all the heat exchangers in line A is presented in the table 6-1. Stream information including temperatures, duties and heat exchangers areas are in the table. In addition the exact name and location of each heat exchanger can be found in the table. For more information regarding the flow rates of each stream, consult appendix 2-1 and 2-2 for chapter 5 under the title of guidelines for constraint analysis in a Kraft process.

Table 6-1: List of Existing heat exchanger - Line A

Location	Name	Area (m^2)	Q (MW)	Hot stream	T_{Hin} (°C)	T_{Hout} (°C)	T_{Cin} (°C)	T_{Cout} (°C)
Digester	Blow Cooler	147	2.79	CBL	95.7	90.0	56.8	80.0
Digester	Flashed Steam Condenser	147	6.46	Flashed steam	99.0	94.9	56.8	70.0
Evaporators	Surface Condenser 1	2916	32.72	Vacuum vapor	59.4	58.4	2.0	59.3
	Surface Condenser 2	241	4.41	Vacuum vapor	78.8	75.8	59.3	67.0
Water Prod	Direct Condenser	N/A	0.00	N/A	N/A	N/A	N/A	N/A
Water Prod	Bleach Heater	50	7.37	LP steam	170	169	68.8	80.0
water prod	Brown Heater	22	3.24	LP steam	170	169	68.8	80.0
Recaust.	Green Liquor Cooler	138	8.53	Green liquor	154	95.0	2.0	80.0
Recaust.	DVSE	746	2.59	Classifier stack	100	98.0	2.0	90.0
Evaporators	Black Liquor Heater	499	17.24	Evap. vapor	91.5	88.5	60.4	60.5
Digester	Cooking Liquor Heater 1	30	2.30	MP steam	202	201	145.6	154.0
Digester	Cooking Liquor Heater 2	43	3.54	MP steam	202	201	138.3	151.8
Steam Plant	Air Heater	281	3.16	LP steam	170	169	45.0	80.0
Machine	Glycol heater	725	1.08	Hot Air	93.4	73.8	61.2	70.0
Machine	Air Heater	204	1.08	Glycol	70.0	61.2	-8.0	25.0
Machine	Air-Air Heater	981	2.17	Dryer Air	132	93.4	25.0	90.0

6.2.2 Existing heat exchanger network – Line B

The heat exchanger network for line B is separated into process network and water network as well. The process network includes all the heat exchangers that do not interact with the heating of water streams. These heat exchangers are:

- Boiler air heater
- Lower and upper cooking liquor heaters
- Glycol loop heaters

The water network includes all the heat exchangers responsible for heating water streams before sending it to the different consumers in the process. Figure 6-2 includes the heat exchangers in the process network and water network.

Figure 6-2: Existing heat exchanger network - Line B

A list of all the heat exchangers in line B process and water network is presented in table 6-2. Stream information including temperatures, duties and areas of heat exchangers area is in the table. In addition the exact name and location of each heat exchanger is in the table.

Table 6-2: List of Existing heat exchangers - Line B

Location	Name	Area (m^2)	Q (MW)	Hot stream	T_{Hin} (°C)	T_{Hout} (°C)	T_{Cin} (°C)	T_{Cout} (°C)
Digester	Cold Blow Cooler	339	13.46	CBL	101	70.0	2.0	60.0
Digester	FSC	46	2.86	FS	100	98.0	50.0	55.0
Pulp Machine	Shower Water Heater	45	1.57	MP Condensate	146.6	65.0	50.0	65.0
Evaporators	Surface Condenser	1181	40.87	vapour	56.7	56.7	2.0	47.0
Water Prod	Direct Condenser	28	4.40	LP Steam	170	169.0	49.7	84.3
Chem Prep	ICC	67	2.69	CLO_2	101	52.6	2.0	50.0
Chem Prep	Surface Condenser	0	0.01	CLO_2	81.3	19.6	2.0	50.0
Recaust	GLC	160	6.07	GL	115	94.0	2.0	95.0
Recaust	Dust Vent Scrubber Exchanger	147	0.52	Classifier vent	100.4	98.1	2.0	90.0
Water Prod	Bleach Heater	17	2.82	LP Steam	180.	179.0	60.0	65.0
Chem Prep	White liquor Cooler	1	0.02	white liquor	98.0	30.0	2.0	20.0
Digester	Upper Cooking Liquor Heater	979	6.11	MP Steam	212	211.0	170	187
Digester	Lower Cooking Liquor Heater	1383	10.29	MP Steam	212	211.0	154	187
Steam Plant	Air Heater	653	7.59	LP Steam	180	179.0	35.0	80.0
Machine	Glycol heater	1238	1.08	Air	93.7	74.8	70.9	75.0
Machine	Air Heater	178	1.08	Glycol	75.0	70.9	-8.0	25.0
Machine	Air-Air Heater	986	2.16	Exhaust	131	93.7	25.0	90.0

6.3 Evaluation of energy violations in the existing network

Two types of energy violations in heat exchangers have been identified. These violations are cross pinch violations and “Criss-cross violations”. They occur due to the inefficient use of energy between hot streams and cold streams in a process.

Cross pinch violations occur when hot streams above the pinch temperature transfer energy to cold streams below the pinch temperature. This inefficient heat transfer will lead to a higher heating requirement above the pinch and higher cooling requirement below the pinch. The cross pinch violation is evaluated quantitatively using Aspen energy analyzer® for line A and line B heat exchanger networks.

Criss-cross violations occurs when a hot stream at a very high temperature level exchanges heat with a cold stream at a low temperature level. Both hot and cold streams could be above or below the pinch. This violation is identified by plotting a temperature level graph for all of the hot streams and cold streams in the heat exchanger network. This is a qualitative tool to help redesign the network in a more efficient way.

The cross pinch and the “Criss-cross violations” for line A and line B are presented and discussed in the following section.

6.3.1 Energy violations - line A

Cross pinch violations

Table 6-3 includes the heat exchangers with cross pinch violations. Green liquor cooler and dust vent scrubber in the recausticizing have the largest cross pinch violations. In addition, flash steam condensers in the digester and air heater in the steam plant have relatively lower yet significant violations. These violations will be eliminated in order to liberate some energy to be used as a heating source in different areas of the process.

Table 6-3: Cross pinch violations - Line A

location	name	Hot Stream	Cold Stream	Energy load (MW)	Cross pinch (MW)
Digester	Blow Cooler	Weak BL	WW	2.79	0.38
Digester	FS Condenser	FS	WW	6.46	1.57
Evaporators	SC 2	Vapor	FW	4.41	0.40
Recaust.	GLC	GL	FW	8.53	6.34
Recaust.	DVSE	Vent Gases	FW	2.59	1.71
Steam Plant	Air Heater	LP Steam	Air	3.16	1.36
Machine	Air-Air Heater	Exhaust Air	Air	2.17	1.17

Criss-cross violations

The most apparent criss cross violations occur in the bleach heater and the brown heater. LP steam at 170 °C is used to heat water from 57 °C to 80 °C. In addition, some heat exchangers such as the green liquor cooler, dust vent scrubber and the boiler air heater have both types of violations. In figure 6-3, the X axis represents the temperature while the Y axis represents the streams used in heat exchangers. The red line s the hot streams of the process while the blues lines represent the cold streams of the process. The length of the stream represents the temperature level across the existing heat exchangers.

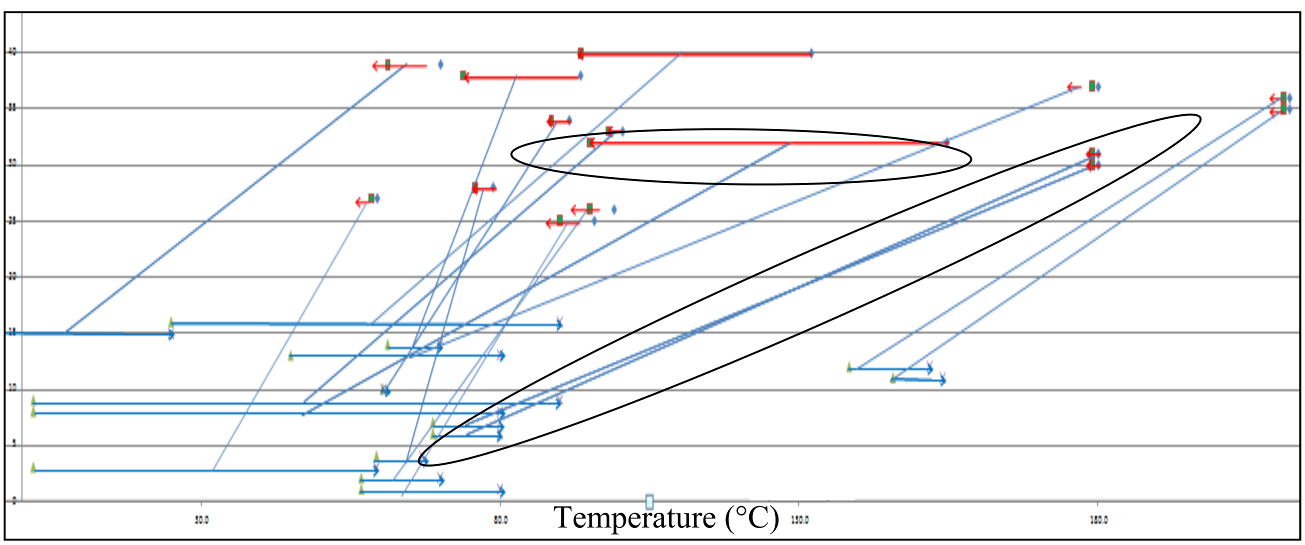


Figure 6-3: Criss cross violations chart - Line A

6.3.2 Energy violations - line B

Cross pinch violations

Table 6-4 includes the heat exchangers with cross pinch violations. Cold blow cooler in the digester and green liquor cooler in the recausticizing have the largest cross pinch violations. In addition, the indirect contact condenser in the chemical preparation and the air heater in the steam plant have lower yet significant violations. Elimination of the violations and the release of the energy to be used in energy saving projects will result in lower minimum heating requirement.

Table 6-4: Cross pinch violations - Line B

location	name	Hot Stream	Cold Stream	Energy load (MW)	Cross pinch (MW)
Digester	Blow Cooler	Weak BL	FW	13.46	10.38
Chem Prep.	ICC	R8 Products	FW	2.69	2.27
Chem Prep.	SC	CLO_2	FW	0.005	0.002
Recaust.	GL Cooler	GL	FW	6.07	2.92
Recaust.	DVSE	Vent Gases	FW	0.52	0.27
Chem Prep.	White liquor Cooler	Weak Liquor	FW	0.018	0.012
Steam Plant	Air Heater	LP Steam	Air	7.59	1.97
Machine	Air Heater	Glycol	Air	1.08	1.08
Machine	Air-Air Heater	Exhaust Air	Air	2.16	0.72

Criss-cross violations

In a similar manner to line A, the Criss-cross chart was drawn and is presented in figure 6-4. The findings were a bit different from line A heat exchangers. The most apparent Criss-cross violations occur in the direct flashed steam condenser in the digester, water condenser in the water production, and the bleach heater. In the bleach heater, LP steam at 170 °C is used to heat water from 57 °C to 80 °C. Some heat exchangers like the boiler air heater have both cross pinch violations of 1.79 MW and criss cross violation whereby LP steam is used to heat boiler air from 35 °C to 80 °C. The chart was built in a similar manner to line A: the X axis represents the temperature level across the streams while the Y axis represents the individual streams. The red line represents the hot streams of the process while the blues lines represent the cold streams of the process.

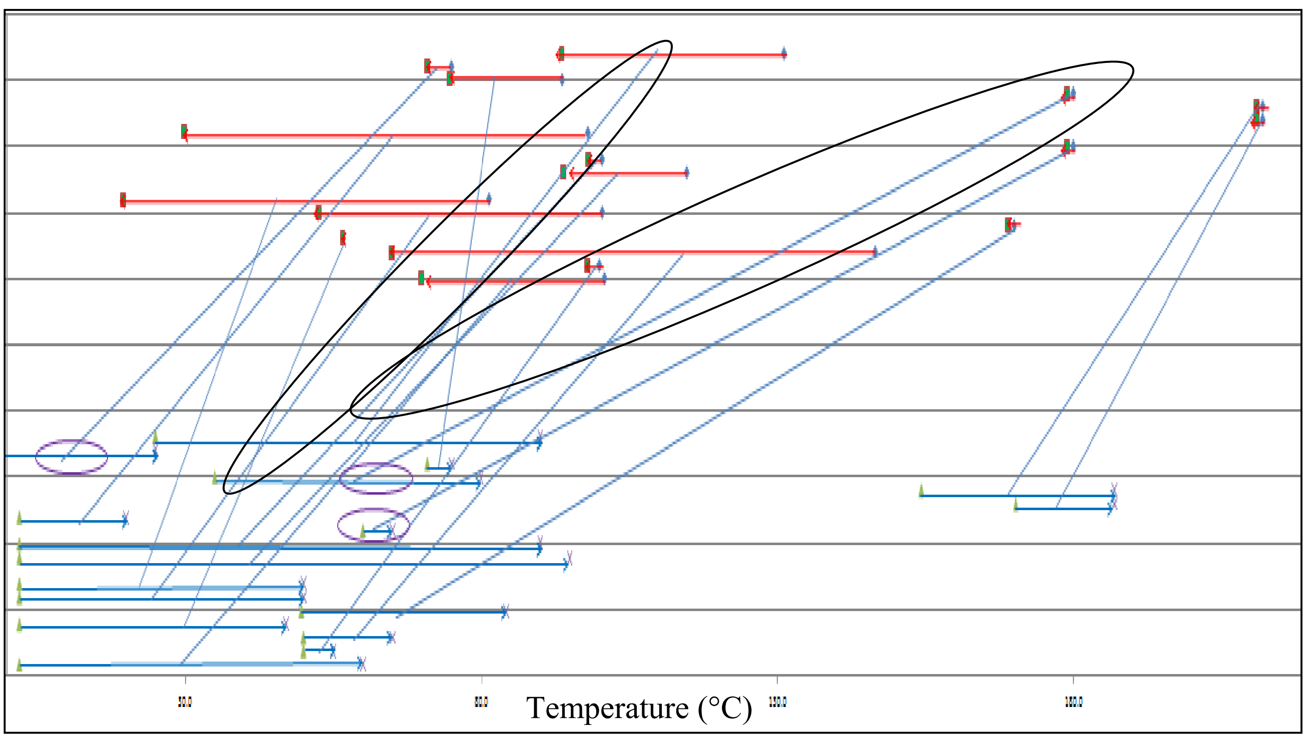


Figure 6-4: Criss cross violations chart - Line B

6.4 Building the new heat exchanger networks

The final step is to implement the energy saving projects into the existing heat exchanger network. At this point of the analysis, the cold streams in the energy saving projects are known while the hot streams are unknown and need to be determined. The tricky task is to choose the hot streams that are a perfect match for the cold stream in each project. There are programs such as Aspen energy analyzer® that produces an automated heat exchanger networks based on the minimization of area but the resulting network is not practical or realistic. Therefore a manual approach to build the heat exchanger network is adopted and it consists of the following steps:

a. Targeting the existing heat exchangers

The heat exchangers with energy violations are addressed at this point. The heat exchangers mentioned in step 2 with high criss-cross violations or pinch violations are targeted and the connection between the hot stream and cold stream is broken thus liberating energy to be used in satisfy the demand of the cold streams in the new energy saving projects.

b. Listing the available streams

Available hot streams such as effluents stack gases, and liberated streams from energy violating heat exchangers are listed. In addition, cold streams that are part of the energy saving projects are listed. Information such as temperature, flow rates and specific heat capacities are required to identify the location of the streams relative to the pinch point. Therefore the streams are split into three main groups; below pinch, above pinch, and below and above pinch. These tables are found in appendix 3-1.

c. Following the rules of pinch

The rules of the pinch technique should be adopted when finding matching streams. These rules can be found in the sources mentioned in the pinch technique literature review section. In general, no cooling utility should be used above the pinch and no heating utility should be used below the pinch. In addition, below the pinch mCp hot streams > mCp cold streams and above the pinch mCp hot streams < mCp cold streams. mCp is the specific heat capacity multiplied by the mass flow rate (kJ/time.°C)

d. Determining the hot stream options for each project

There could be more than one suitable stream for each project or cold demand and therefore the compatible hot streams are listed for each project. A screening is done in the next step to determine the most suitable stream for each project.

e. Final selection of streams

Screening of the compatible hot streams to determine the most suitable stream for each heat exchanger or energy saving project is the final step. The criteria for screening should include: Area of potential heat exchanger, location of streams, and distances between streams. In addition, one should always try to find scenarios of projects with the least number of heat exchanger and the highest potential of savings with minimum energy violations. Finally, the pinch rules should be kept in mind at every stage of the screening process.

The energy saving projects for line A are proposed and combined to form new heat exchanger networks. Three heat exchanger networks for line A are proposed at three different constraint levels. The projects in the three scenarios are similar but the degree of energy savings varies significantly. The heat exchanger networks are presented in the following manner:

I. Retrofit – low savings
II. Retrofit – medium savings
III. Grassroot – high savings

On the other hand, one heat exchanger networks has been developed for line B which is the retrofit – medium savings. The evaluation of the projects and the heat exchanger network is presented after line A networks.

6.4.1 Line A - Retrofit – Low savings

In the retrofit – low savings constraint level, minor adjustments are made to the existing heat exchanger network by targeting three steam heat exchangers and two steam direct injection points. The direct injection points are located in the deaerator and bleaching steam injection 1. Low pressure steam is replaced by other hot effluents and stack gases in the process. In addition, heat exchangers that are violating the pinch are kept intact. The proposed energy saving projects are presented below:

Project 1 – Bleach Heater:

In the current configuration bleach heater uses low pressure (LP) steam to heat water from hot water tank to bleaching consumers. The temperature of the water increases from 68 °C to the target temperature of 80 °C. The proposed design suggests the replacement of LP steam by recovery boiler 1 stack gases. The steam savings incurred are 7.36 MW. The project schematic is presented in figure 6-5.

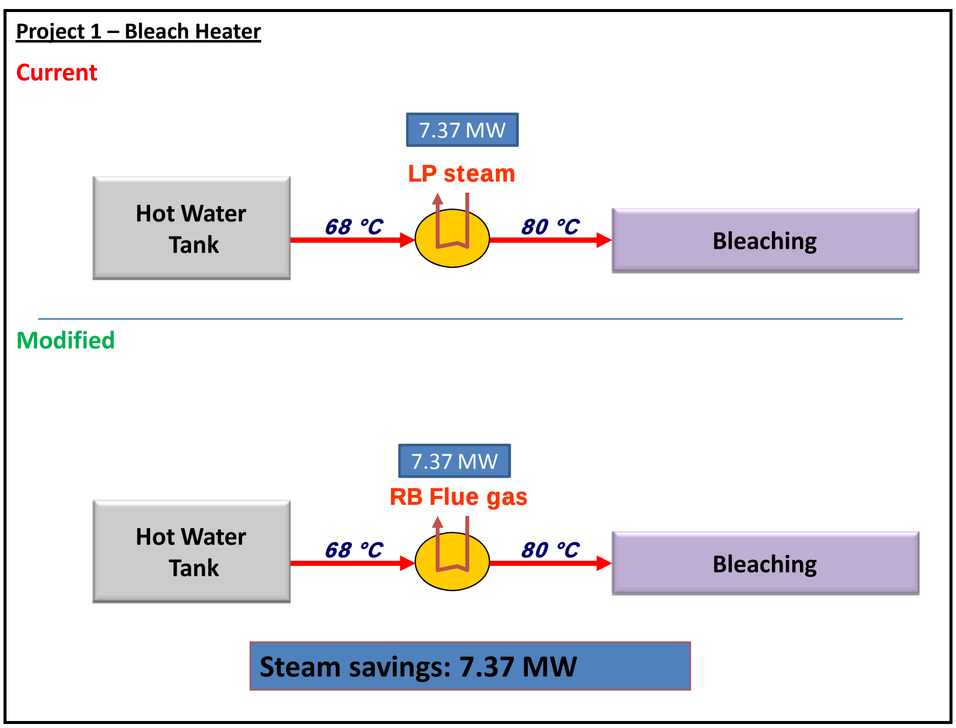


Figure 6-5: Project 1 - Bleach Heater

<u>Project 2 – Brownstock Heater:</u>

In the current configuration brown heater uses low pressure (LP) steam to heat water from hot water tank to brownstock washing consumers. The temperature of the water increases from 68 °C to the target temperature of 80 °C. The proposed design suggests the replacement of LP steam by blowdown from recovery boiler 1 and Limekiln stack gases. The steam savings incurred are 3.24 MW. The project schematic is in figure 6-6.

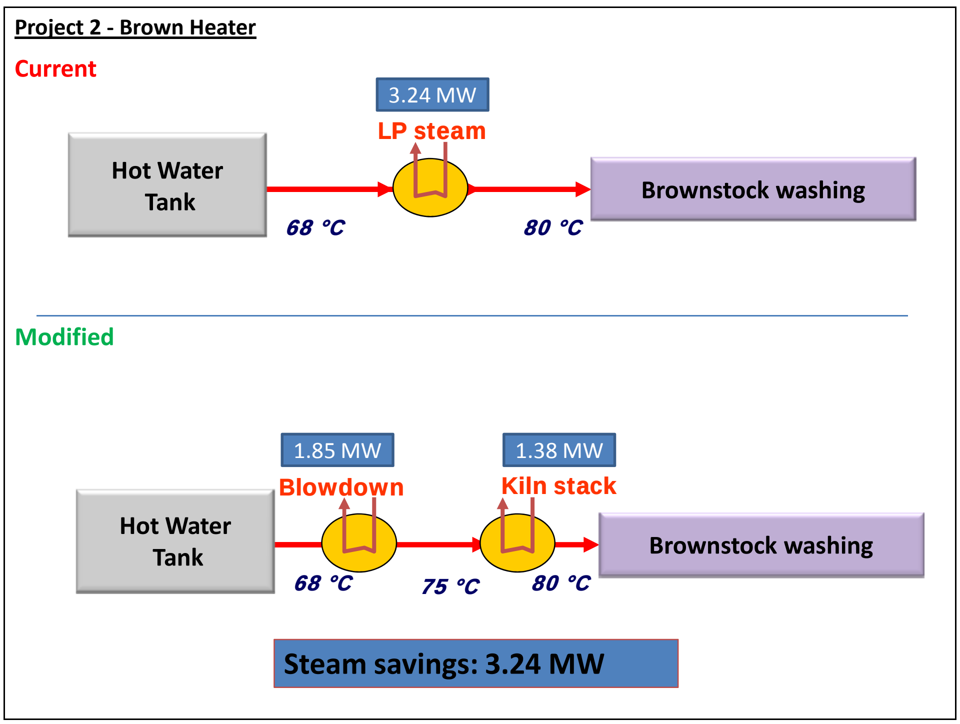


Figure 6-6: Project 2- Brown Heater

Project 3 – Boiler Air Heater:

In the current configuration boiler heater uses low pressure (LP) steam to heat Air entering into the recovery boiler in steam plant. The temperature of the air increases from 45 °C to the target temperature of 80 °C. The proposed design suggests the replacement of LP steam by alkaline effluent below the pinch and recovery boiler stack gas above the pinch. The steam savings incurred are 3.16 MW. The project schematic is presented in figure 6-7.

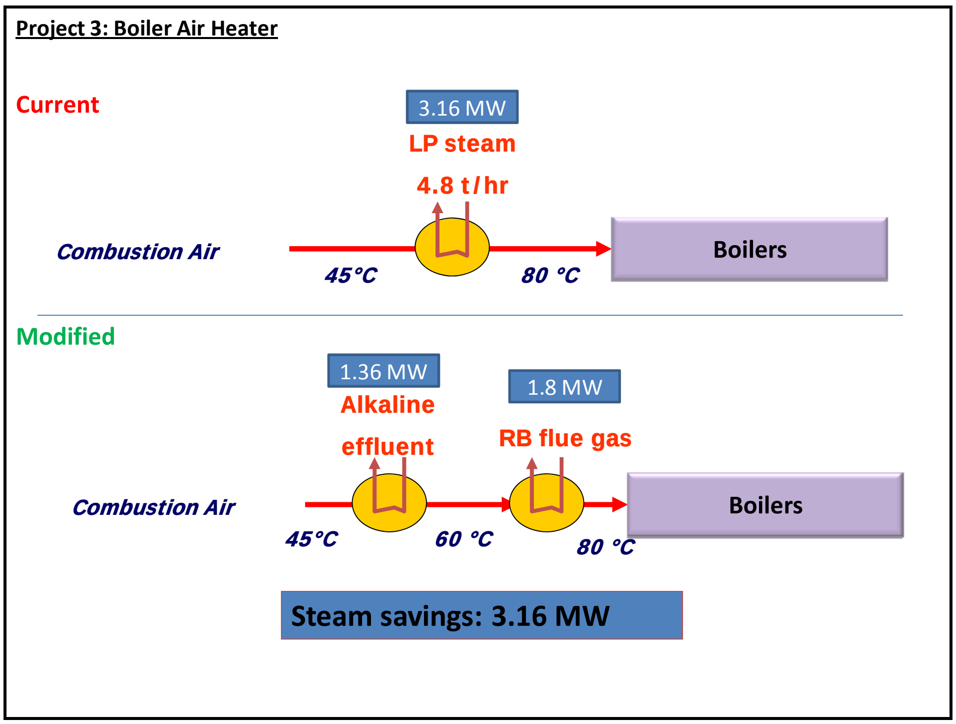


Figure 6-7: Project 3 - Boiler Air Heater

Project 4 – Make up water (Deaerator):

In the current configuration, boiler make up water at 31 °C is mixed with condensate at 102 °C in condensate collection tank resulting in a non isothermal mixing point. Water at 65 °C is sent from the tank into the deaerator to be heated with direct steam. The proposed design focuses on the elimination of the non isothermal mixing in the tank thus reducing the steam injection in the deaerator. This can be achieved by heating the makeup water to 95 °C using alkaline effluent below the pinch and recovery boiler stack gases above the pinch. The steam injected into the deaerator is reduced and a saving of 6 MW is obtained. The project schematic is presented in figure 6-8.

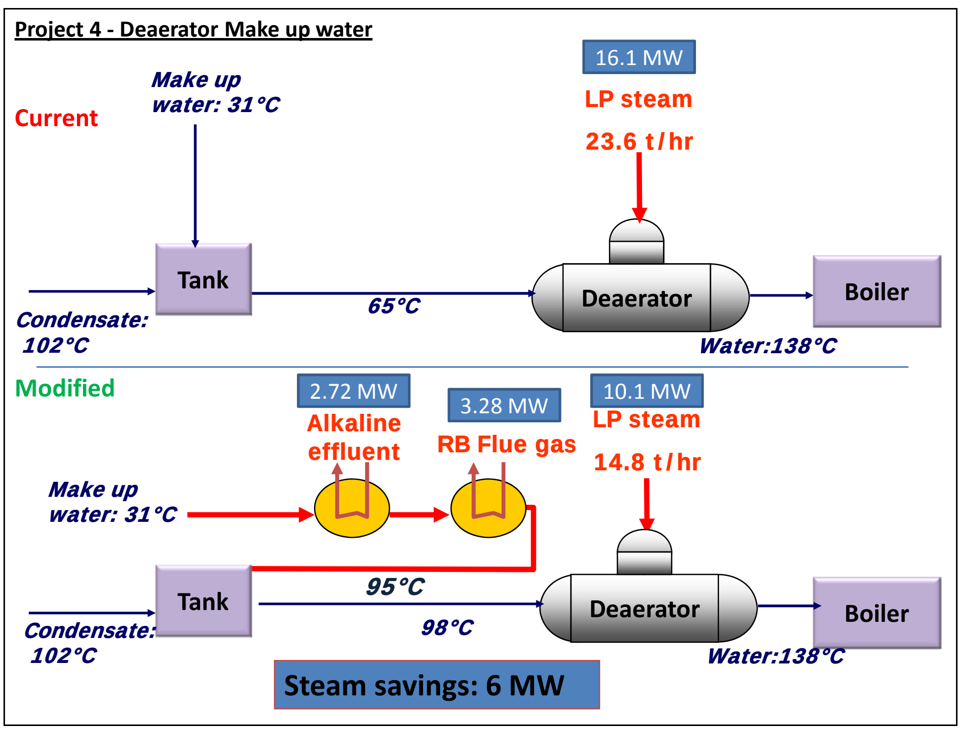


Figure 6-8: Project 4- Deaerator Mae up water

Project 5 – Injection 1 – Washer 15:

In the current configuration, steam is being injected in the bleaching pulp line to heat the pulp to 81 °C. White water is being injected into washer 15 at 57 °C resulting in a non isothermal mixing point in the washer 15. The proposed design suggests the heating of white water to 81 °C using recovery boiler stack before injecting it into washer 15. This will eliminate the non isothermal mixing point and result in steam savings of 1.7 MW. The project schematic is presented in figure 6-9.

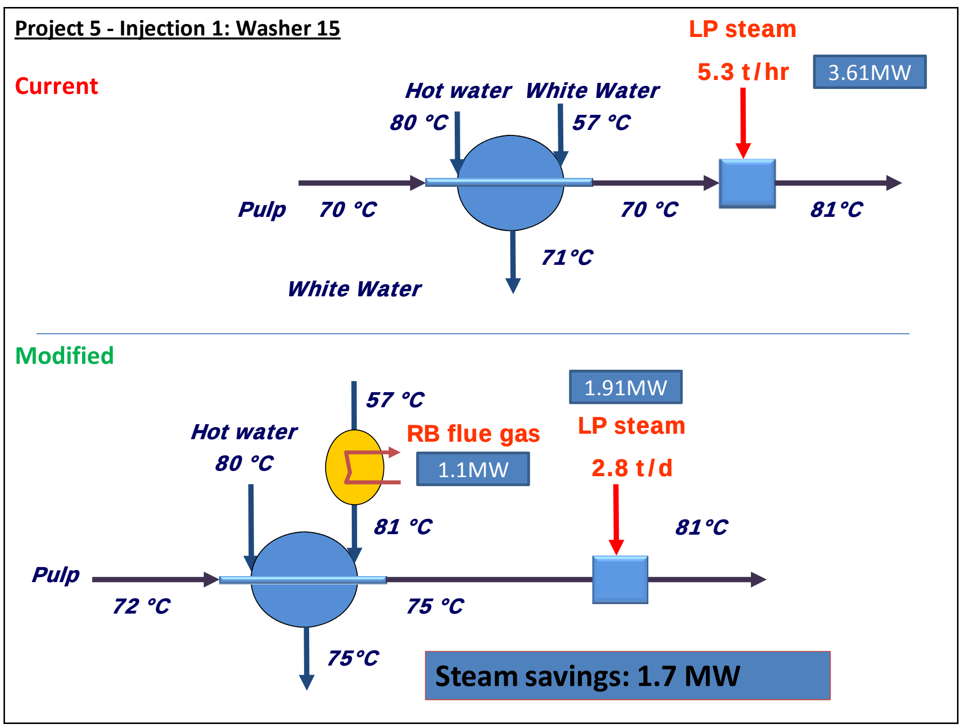


Figure 6-9: Project 5 - Injection 1: Washer 15

6.4.1.1 Summary of energy saving projects – Retrofit low savings

The summary of the energy saving projects are presented in table 6-5. The most promising projects in terms of energy savings are replacing the bleach heater and heating the makeup water. Nonetheless the other projects have savings ranging between 1% and 2 %. The total energy savings based on the compilation of the projects are 21.5 MW or 11% of total steam consumption.

Table 6-5: Summary of energy saving projects - Retrofit low savings

Project	Description	Steam Savings (MW)	Steam Savings (%)
1	Replace bleach heater	7.4	4%
2	Upgrade brown heater	3.2	2%
3	Upgrade Air heater	3.2	2%
4	Heat Make up water	6.0	3%
5	Reduce Bleaching steam injection	1.7	1%
	Total	21.5	11

6.4.2 Line A - Retrofit – Medium savings

In the retrofit – medium savings constraint level, major adjustments are made to the existing heat exchanger network by retrofitting the cross pinch heat exchangers and eliminating or reducing indirect steam consumption in heaters. In addition, Non isothermal mixing points are eliminated by focusing on heating water and liquor streams prior to mixing; this evidently will lead to the reduction of direct injection steam. The direct injection points are located in the deaerator and bleaching steam injection points 1-4. Low pressure steam is replaced by other liberated process streams, effluents and stack gases. The proposed potential energy saving projects are presented:

<u>Project 1 – Bleach Heater</u>

In the current configuration Green liquor is being cooled in green liquor cooler from 155 °C to 95 °C using fresh water. The fresh water is heated from 2 °C to 80 °C resulting in a cross pinch violation of 6.34 MW. In the Dust vent scrubber heat exchanger water is being heated to 90 °C while violating the pinch by 1.71 MW. Finally, bleach heater uses LP steam to heat water from hot water tank to bleaching consumers. The temperature of the water increases from 68 °C to the target temperature of 80 °C. The proposed design suggests the elimination of pinch violations in both heat exchangers and the release of green liquor to be used in the bleach heater instead of LP steam. The steam savings incurred are 7.36 MW. The project is presented in figure 6-10.

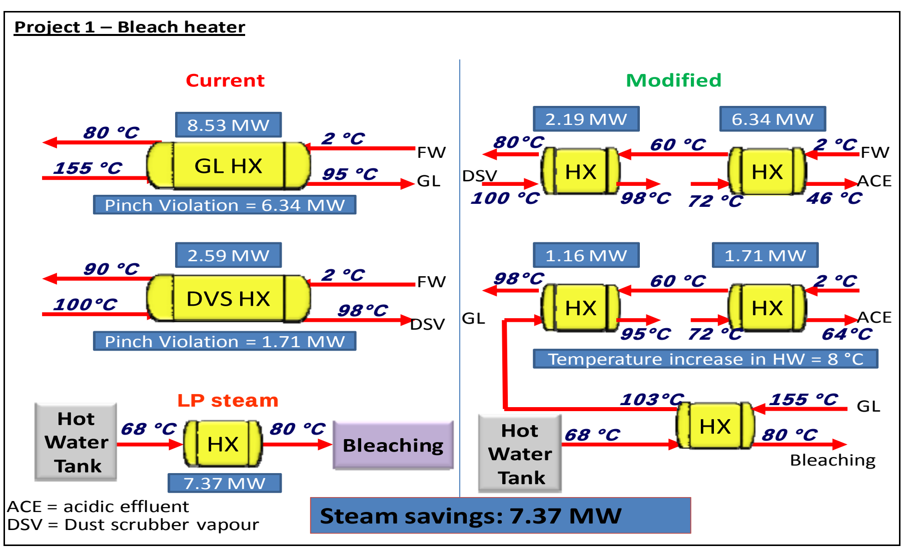


Figure 6-10: Project 1 - Bleach Heater

Project 2 – Brownstock Heater:

In the current configuration flashed steam from flash tank 2 in digester department is used to heat warm water to 70 °C with a pinch violation of 1.57 MW. The heated water is then sent to hot water tank. In addition, brown heater uses low pressure (LP) steam to heat water from hot water tank to brownstock washing consumers. The temperature of the water increases from 68 °C to the target temperature of 80 °C. The potential project involves the addition release of energy due to the elimination of cross pinch in the flashed steam condenser and the use of blowdown as an energy source. By using alkaline below the pinch to heat warm water, 1.57 MW of flashed steam is used to heat water from hot water tank to 74 °C before it is heated by blowdown to the target temperature of 80 °C. The steam savings incurred are 3.24 MW. The project schematic is presented in figure 6-10.

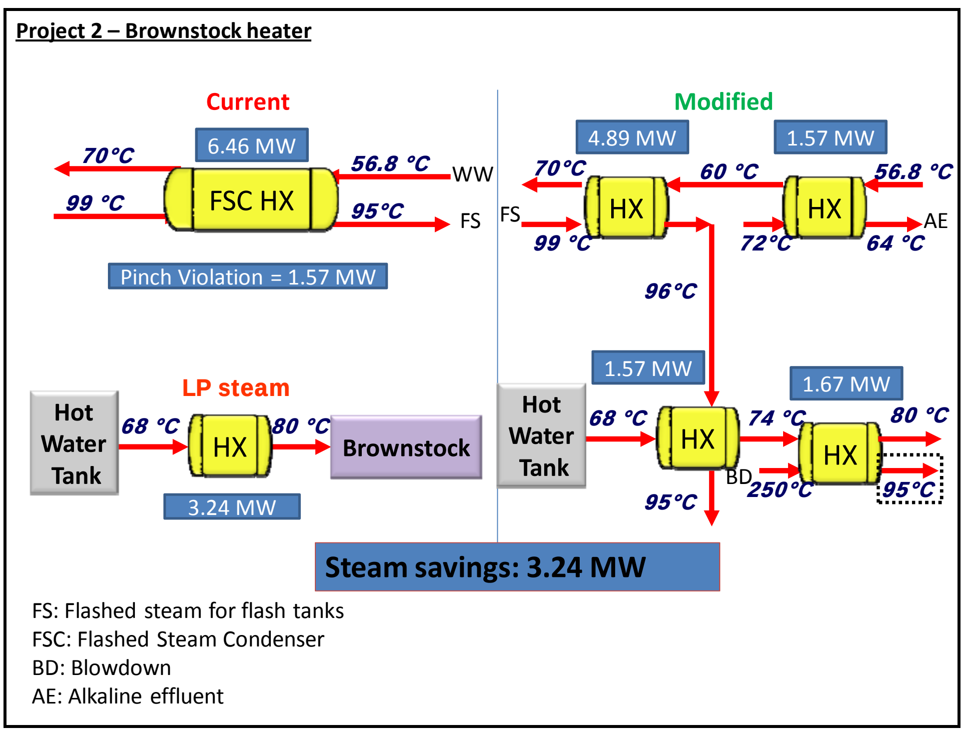


Figure 6-11: Project 2 - Brownstock heater

Project 3 – Boiler Air Heater:

In the current configuration boiler heater uses low pressure (LP) steam to heat Air entering into the recovery boiler in steam plant. The temperature of the air increases from 45 °C to the target temperature of 80 °C. The proposed design suggests the replacement of LP steam by alkaline effluent below the pinch and recovery boiler stack gas above the pinch. The steam savings incurred are 3.16 MW. The project schematic is presented in figure 6-12.

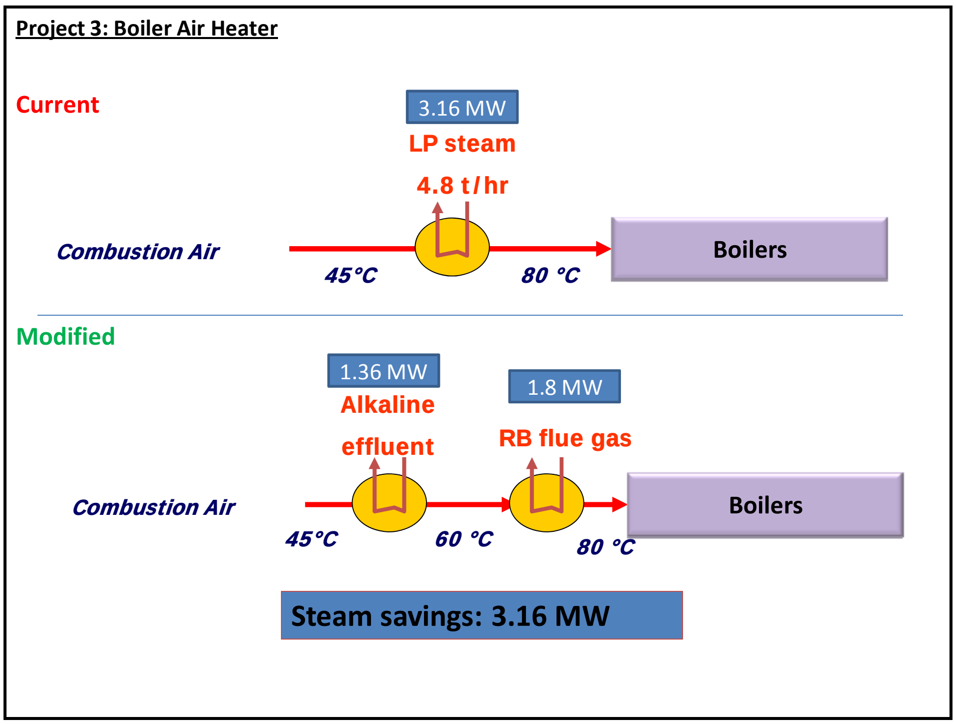


Figure 6-12: Project 3 - Boiler Air Heater

Project 4 – Make up water (Deaerator) :

In the current configuration, boiler make up water at 31 °C is mixed with condensate at 102 °C in condensate collection tank resulting in a non isothermal mixing point. Water at 65 °C is sent from the tank into the deaerator to be heated with direct steam. The proposed design focuses on the elimination of the non isothermal mixing in the tank thus reducing the steam injection in the deaerator. This can be achieved by heating the makeup water to 95 °C using alkaline effluent below the pinch and recovery boiler stack gases above the pinch. The steam injected into the deaerator is reduced and a saving of 6 MW is obtained. The project schematic is presented in figure 6-13.

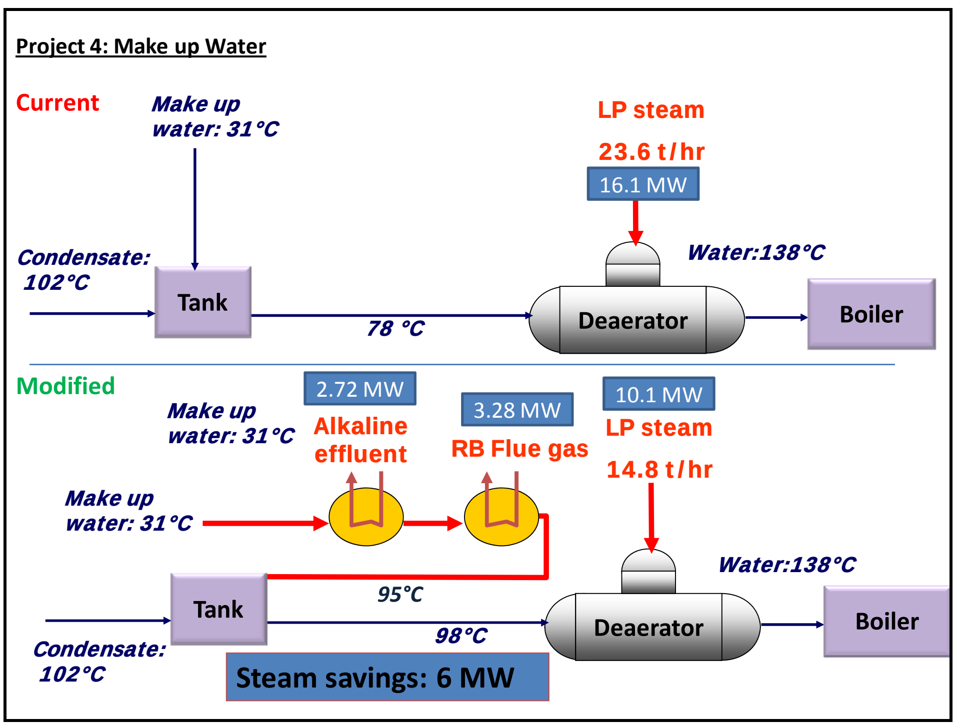


Figure 6-13: Project 4 - Make up Water

Project 5 – Injection 1 – Washer 15:

In the current configuration, steam is being injected in the bleaching pulp line to heat the pulp to 81 °C. White water is being injected into washer 15 at 57 °C resulting in a non isothermal mixing point in the washer 15. The proposed design suggests the heating of white water to 81 °C using recovery boiler stack before injecting it into washer 15. This will eliminate the non isothermal mixing point and result in steam savings of 1.7 MW. One thing to note is that the heat exchanger used in this project is the same as the heat exchanger used in project 8 to heat white water. The heat exchanger was separated to make the projects more clear. The project schematic is presented in figure 6-14.

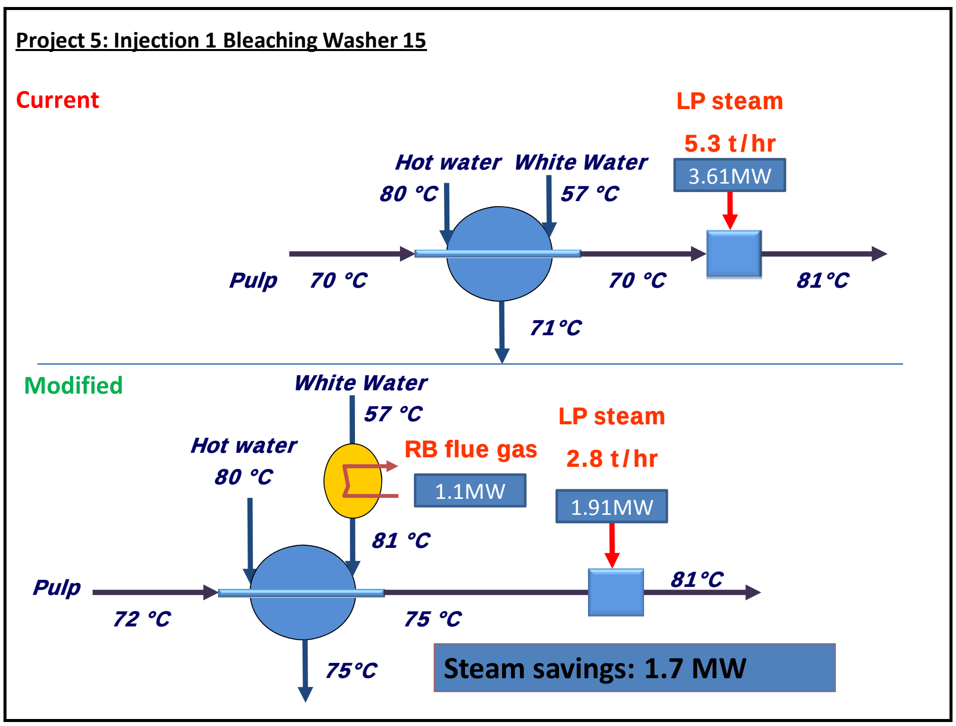


Figure 6-14: Project 5 - Injection 1 Bleaching Washer 15

<u>Project 6 – Injection 2 – Washer 35:</u>

In the current configuration, Fresh water is being injected into the pulp line resulting in a non isothermal mixing point of with a temperature drop of 2 °C. Washing liquors area injected into the washer at a temperature between 67-70 °C. Finally, steam is being injected in the bleaching pulp line to heat the pulp to 84 °C. The proposed design suggests the heating of the fresh water, washing liquor 1 and 2 to 60 °C, 81 °C and 81 °C respectively using acidic effluent and recovery boiler stack gases. By doing so, the temperature of the pulp exiting the washer increases by 9 °C which will result in a steam saving of 2.7 MW. One thing to note is that the heat exchanger used to heat liquor at 70 °C is the same as the heat exchanger used in project 7 to heat liquor at 70 °C. The heat exchanger was separated to make the projects more clear. The project schematic is presented in figure 6-15.

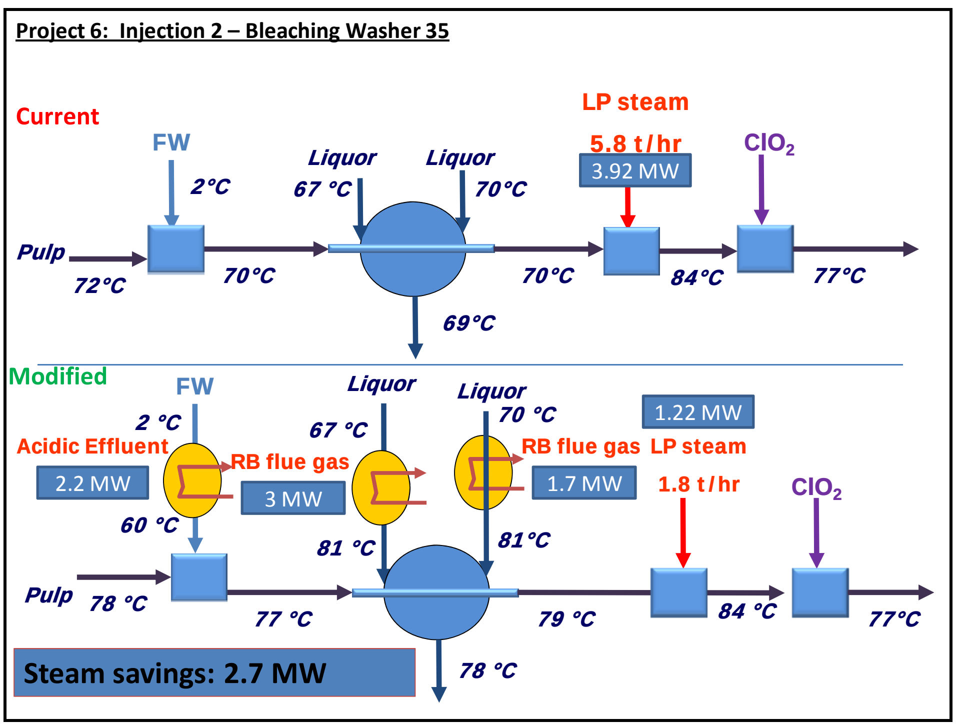


Figure 6-15: Project 6 - Injection 2 – Bleaching Washer 35

Project 7 – Injection 3 – Washer 45:

In the current configuration, Fresh water is being injected into the pulp line resulting in a non isothermal mixing point of with a temperature drop of 1 °C. Washing liquors area injected into the washer at a temperature between 64-70 °C. Finally, steam is being injected in the bleaching pulp line to heat the pulp to 79 °C. The proposed design suggests the heating of the fresh water, washing liquor 1 and 2 to 60 °C, 80 °C and 80 °C respectively using acidic effluent and recovery boiler stack gases. By doing so, the temperature of the pulp exiting the washer increases by 10 °C which will result in a steam saving of 2.9 MW. One thing to note is that the heat exchanger used to heat liquor to 70 °C is the same as the heat exchanger used in project 6 to heat liquor at 70 °C. In addition, the heat exchanger used to heat liquor at 64 °C is the same as the heat exchanger used in project 8 to heat liquor at 64 °C. The project schematic is presented in figure 6-16.

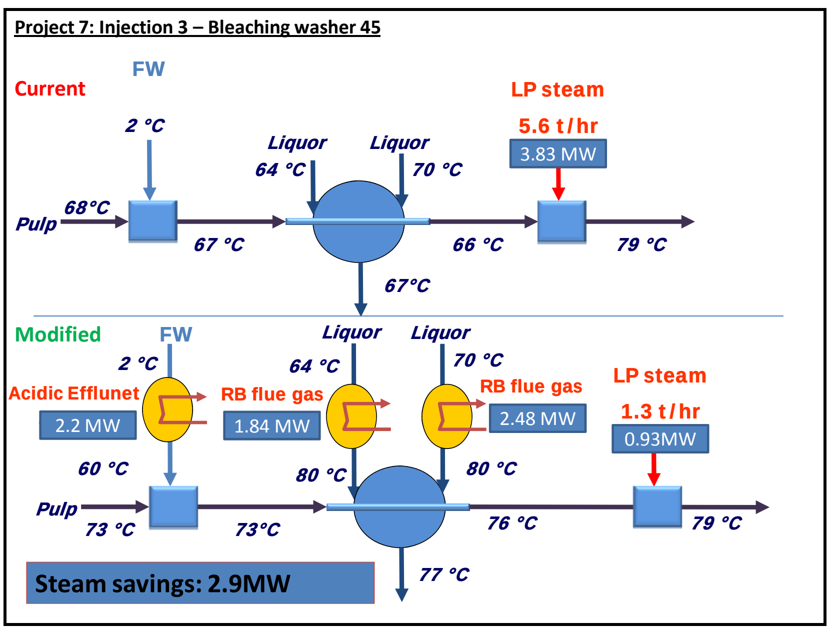


Figure 6-16: Project 7 – Injection 3 – Washer 45

Project 8 – Injection 4 – Washer 55:

In the current configuration, Washing liquor and white water are injected into the washer at a temperature between 57 - 64 °C. This results in a non isothermal mixing and a temperature drop in the pulp line of about 7 °C. Steam is being directly injected in the bleaching pulp line to heat the pulp to 79 °C. The proposed design suggests heating, washing liquor 1 and 2 to 60 °C and 80 °C respectively using recovery boiler stack gases. By doing so, the temperature of the pulp exiting the washer increases by 5 °C which will result in a steam saving of 1.6 MW. One thing to note is that the heat exchanger used to heat white water at 57 °C is the same as the heat exchanger used in project 5 to heat white water. In addition, the heat exchanger used to heat liquor at 64 °C is the same as the heat exchanger used in project 7 to heat liquor at 64 °C. The project schematic is presented in figure 6-17.

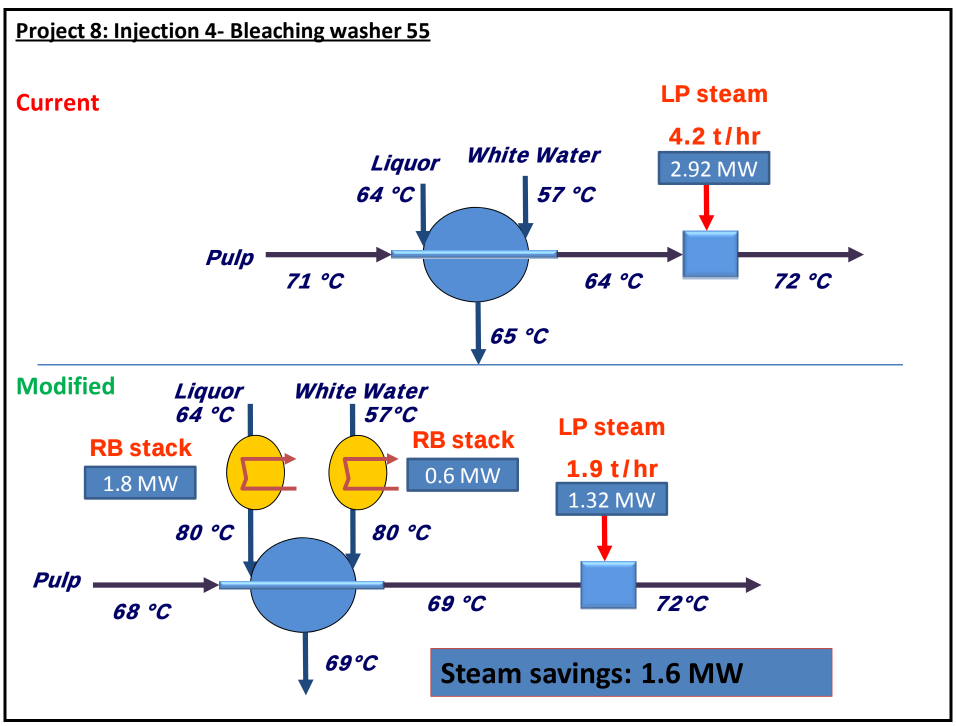


Figure 6-17: Project 8 – Injection 4 – Washer 55

6.4.2.1 Summary of potential energy savings projects – Retrofit medium savings

The summary of the energy saving projects are presented in table 6-6. The most promising projects in terms of energy savings are replacing the bleach heater and heating the makeup water. Nonetheless the other projects have savings ranging between 1% and 2 %. The total energy savings based on the compilation of the projects are 28.7 MW or 15 % of total steam consumption. It is evident that the steam savings increased due to the release of hot streams that were crossing the pinch in the current heat exchanger network. The number of energy savings projects increased because of the increase in the availability of hot streams.

Table 6-6: Potential energy savings - Retrofit medium savings

#	Project name	Steam Savings (MW)	Steam Savings (%)
1	Bleach heater	7.37	3.9
2	Brown heater	3.24	1.7
3	Boiler Air Heater	3.16	1.7
4	Deaerator (make up water)	6.00	3.0
5	Injection 1 (Washer 15)	1.70	0.9
6	Injection 2 (Washer 35)	2.70	1.7
7	Injection 3 (Washer 45)	2.94	1.5
8	Injection 4 (Washer 55)	1.60	0.6
Total		28.7	15

6.4.3 Line A – Grassroot - High savings

In the Grassroot - high savings constraint level, the current heat exchanger network undergoes complete restructuring in order to maximize the savings. This is done by making a completely new design without any constraints due to previous connections in the various heat exchangers. The focus is mainly on finding the best hot stream to satisfy the cold water streams demand. The savings are obtained by eliminating or reducing indirect steam consumption in heaters. In addition, Non isothermal mixing points are eliminated by focusing on heating water and liquor streams prior to mixing; this evidently has will lead to the reduction of direct injection steam. The direct injection points are located in the deaerator and bleaching steam injection points 1-4. The proposed potential energy saving projects are presented below.

<u>Project 1 and 2 – Warm water to Washing and Bleaching</u>

In this project vacuum vapor from evaporators is used to heat fresh water from 2 °C to 56.8 °C. The water is then sent to bleaching and washing departments. Both water streams are heated in the same heat exchanger. The rest of energy in vacuum vapor is used in project 3 and 4 in a single heat exchanger and in project 7 as well. There are no steam savings in this specific project but the energy in vacuum vapor is used in a more efficient way resulting in higher amount of free energy to be used in energy saving projects. The project schematic is presented in figure 6-18.

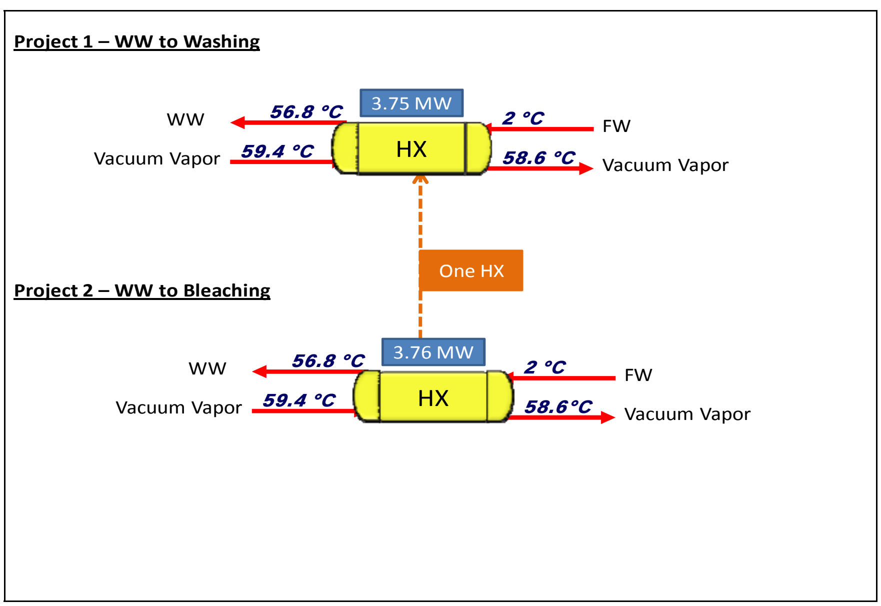


Figure 6-18: Project 1& 2 – WW to Washing and Bleaching

Project 3 – Hot water to bleaching:

In this project hot water to bleaching is heated in three different stages. The first stage corresponds to the use of evaporator vacuum vapor to heat water to a temperature of 54.5 °C. The water entering into the first stage is a mix of water from project 3 as well as project 4. This will lead to a better controllability of the heat exchanger network and will reduce the number of required heat exchangers by one. In the second stage, water is heat to the pinch temperature of 59.5 using alkaline effluent. The water entering into the second stage is a mix of water from project 3 as well as project 4. In the last stage, water is heated to the required temperature of 80 °C using flashed vapor from the second flash tank in the digester department. There are no steam savings in this specific project but the energy is distributed more efficiently. The project schematic is presented in figure 6-19.

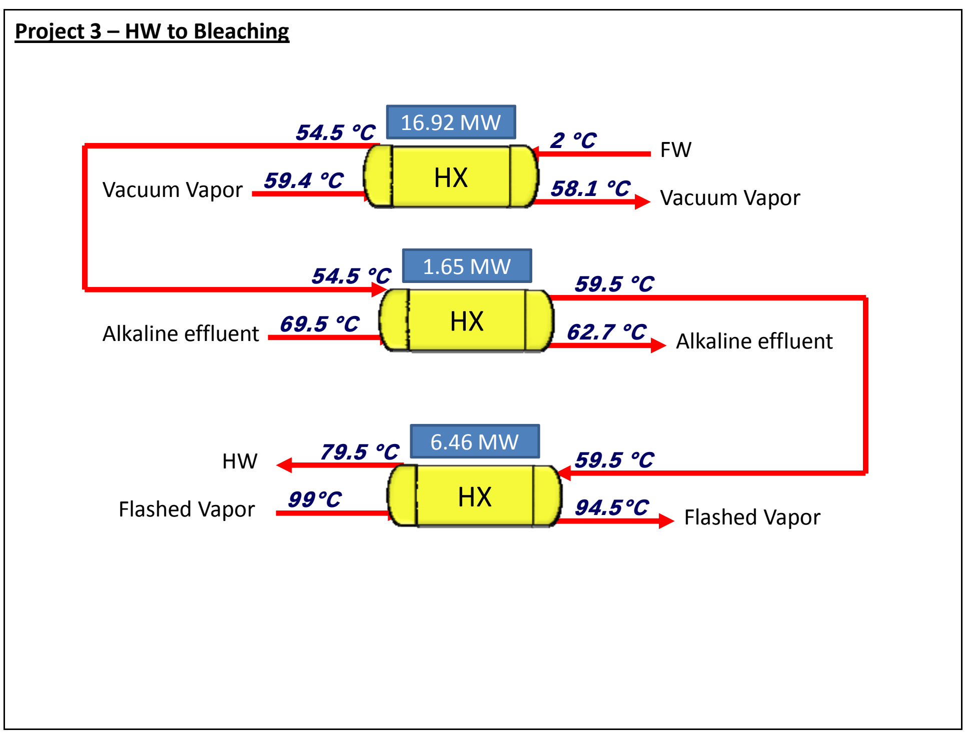


Figure 6-19: Project 3 – Hot water to bleaching

<u>Project 4 – Hot water to Machine:</u>

Project 4 entails the heating of fresh water to produce hot water that is sent to the pulp machine. This project is similar to project 3 where by hot water is heated in three different stages. The first stage corresponds to the use of evaporator vacuum vapor to heat water to a temperature of 54.5 °C. The water entering is a mix of water from project 3 as well as project 4. In the second stage, water is heat to the pinch temperature of 59.5 using alkaline effluent. The water entering into the first stage is a mix of water from project 3 as well as project 4. In the last stage, water is heated to the required temperature of 80 °C using green liquor in the recausticizing department. There are no steam savings in this specific project but the energy is distributed more efficiently. The project schematic is presented in figure 6-20.

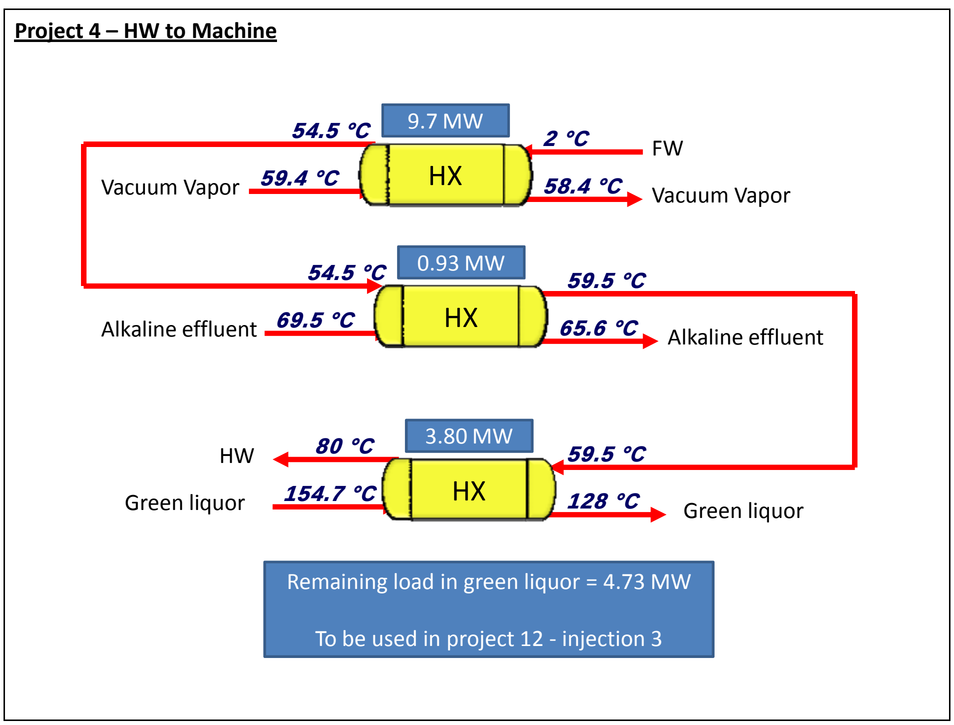


Figure 6-20: Project 4 – Hot water to Machine

<u>Project 5 – Hot water to Recausticizing 1:</u>

In project 5, water to recausticizing is heated in two distinct stages. In the first stage fresh water is heated below the pinch to 59.5 °C using acidic effluent in bleaching. In the second stage water is heated above the pinch to 80 °C using classifier stack gases in the recausticizing. The project schematic is presented in figure 6-21.

<u>Project 6 – Hot water to Recausticizing 2:</u>

In project 5, water to recausticizing is heated in two distinct stages. In the first stage fresh water is heated below the pinch to 59.5 °C using Machine air in the pulp machine. In the second stage water is heated above the pinch to 90 °C using weak black liquor in digester. Heat load of 1.9 MW remains in the weak black liquor and is used in project 8 to heat water to bleaching from line B. The project schematic is presented in figure 6-21.

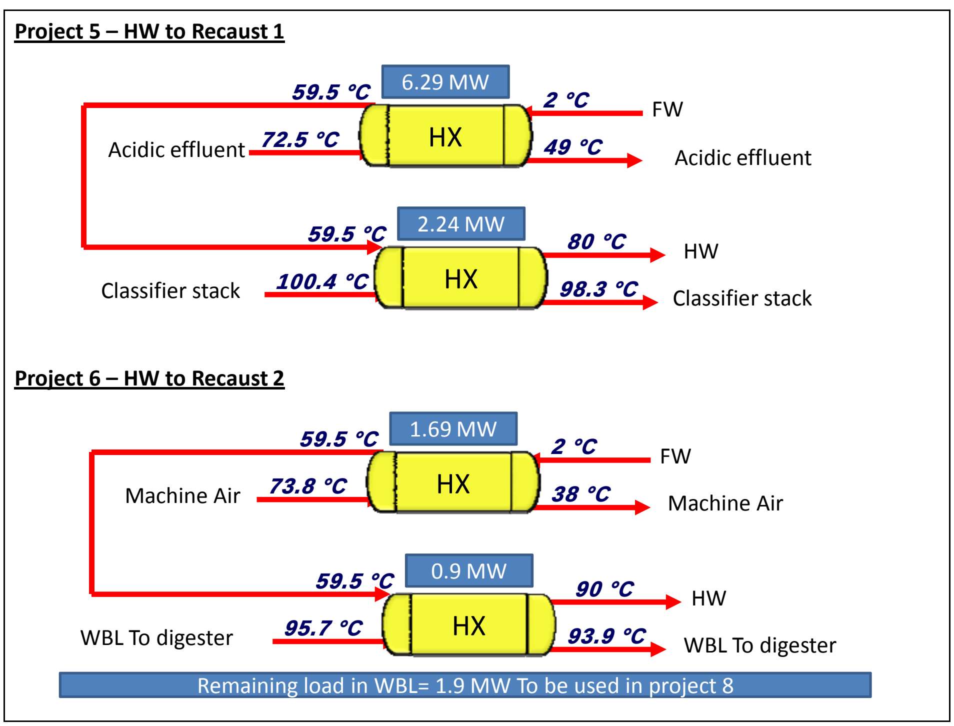


Figure 6-21: Project 5 & 6 – Hot water to Recausticizing 1 & 2

<u>Project 7 – Hot water to Bleaching:</u>

Project 7 consists of three heating stages to heat fresh water to produce hot water that is sent to the Bleaching. In stage one acidic effluent in bleaching is used to heat the water to 59°C.5below the pinch. A small pinch violation is allowed at this point. The second stage corresponds to the use of evaporator vacuum vapor 2 (higher pressure) to heat water to a temperature of 76.3 °C. In the last stage, lime kiln stack gas is used to heat the water to the target temperature of 80 °C .There are no steam savings in this specific project but the energy is distributed more efficiently than the current situation. The project schematic is presented in figure 6-22.

<u>Project 8 – Hot water to Machine:</u>

This project only involves a single stage heating using the remaining energy in weak black liquor after project 6. Warm Water from line B is heated to 80 °C before it is being sent to bleaching. The project schematic is presented in figure 6-23.

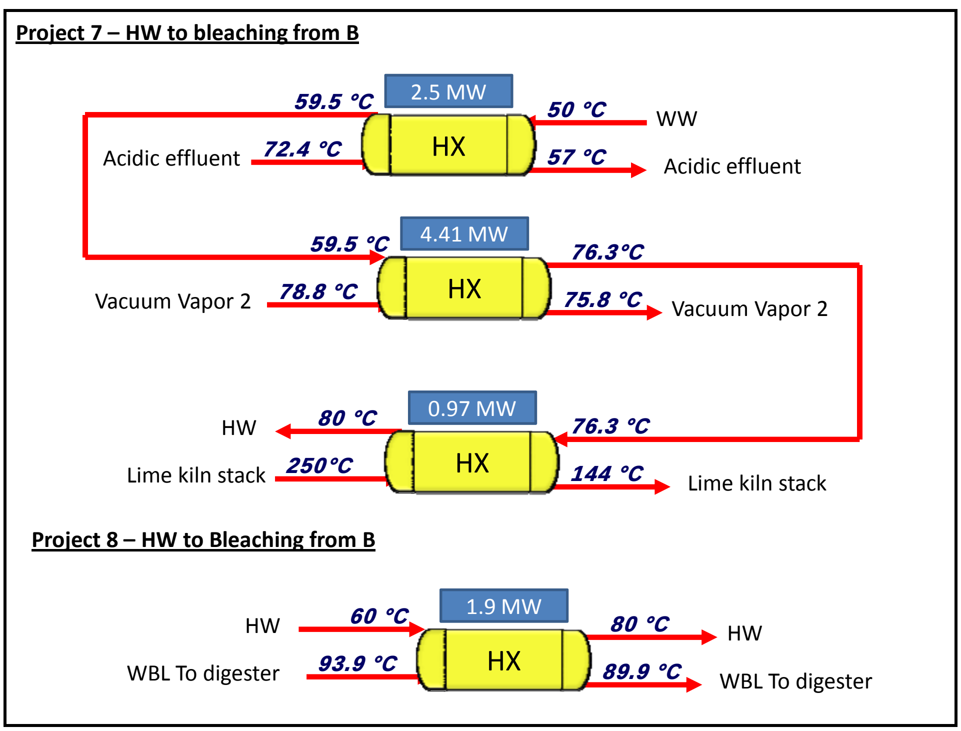


Figure 6-22: Project 7 – Hot water to Bleaching

Project 9 – Make up water (Deaerator):

In the current configuration, boiler make up water at 31 °C is mixed with condensate at 102 °C in condensate collection tank resulting in a non isothermal mixing point. Water at 65 °C is sent from the tank into the deaerator to be heated with direct steam. The proposed design focuses on the elimination of the non isothermal mixing in the tank thus reducing the steam injection in the deaerator. This can be achieved by heating the makeup water to 98 °C using alkaline effluent below the pinch and recovery boiler stack gases above the pinch. The steam injected into the deaerator is reduced and a saving of 6.37 MW is obtained. The project schematic is presented in figure 623.

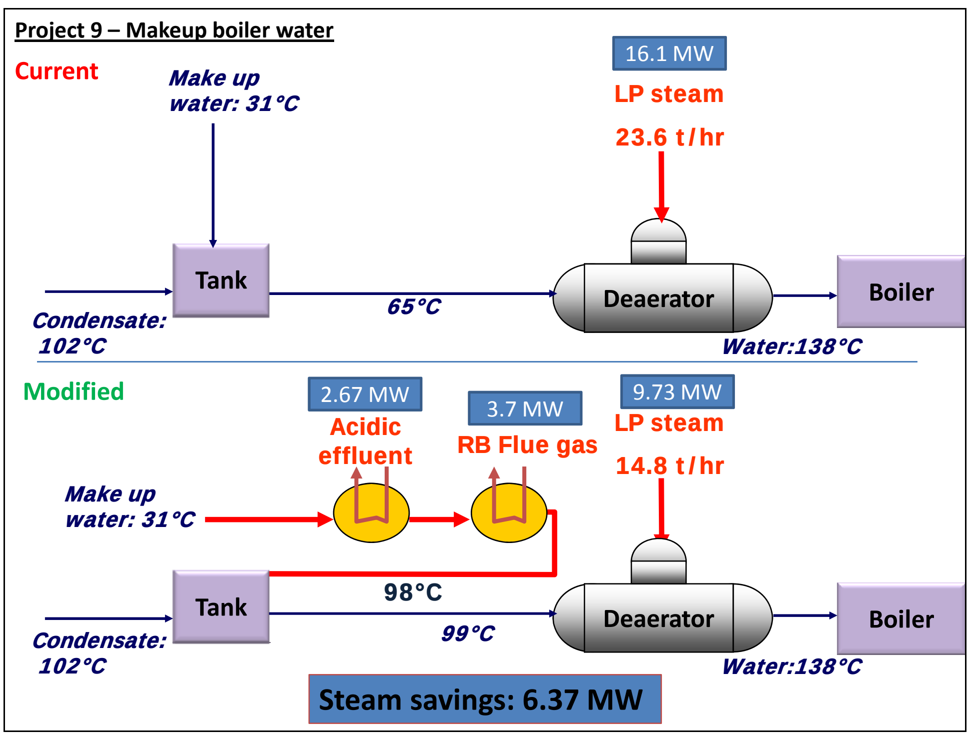


Figure 6-23: Project 9 – Make up water (Deaerator)

Project 10 – Injection 1 – Washer 15:

In the current configuration, steam is being injected in the bleaching pulp line to heat the pulp to 81 °C. White water is being injected into washer 15 at 57 °C resulting in a non isothermal mixing point in the washer 15. The proposed design suggests the heating of white water to 82 °C using recovery boiler stack before injecting it into washer 15. This will eliminate the non isothermal mixing point and result in steam savings of 2.11 MW. One thing to note is that the heat exchanger used in this project is the same as the heat exchanger used in project 13 to heat white water. The heat exchanger was separated to make the projects representation more clear. The project schematic is presented in figure 6-24.

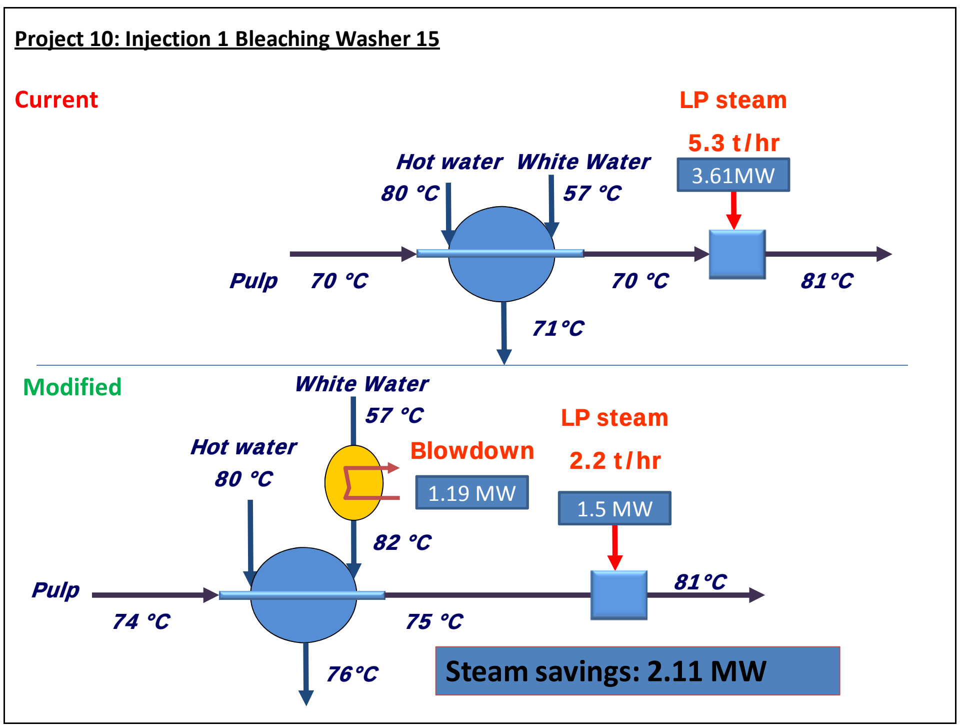


Figure 6-24: Project 10 – Injection 1 – Washer 15

Project 11 – Injection 2 – Washer 35:

In the current configuration, Fresh water is being injected into the pulp line resulting in a non isothermal mixing point of with a temperature drop of 2 °C. Washing liquors area injected into the washer at a temperature between 67-70 °C. Finally, steam is being injected in the bleaching pulp line to heat the pulp to 84 °C. The proposed design suggests the heating of the fresh water, washing liquor 1 and 2 to 60 °C, 88 and 87 °C respectively using acidic effluent and recovery boiler stack gases. By doing so, the temperature of the pulp exiting the washer increases by 9 °C which will result in a steam saving of 3.76 MW. One thing to note is that the heat exchanger used to heat liquor at 70 °C is the same as the heat exchanger used in project 12 to heat liquor at 70 °C. The project schematic is presented in figure 6-25.

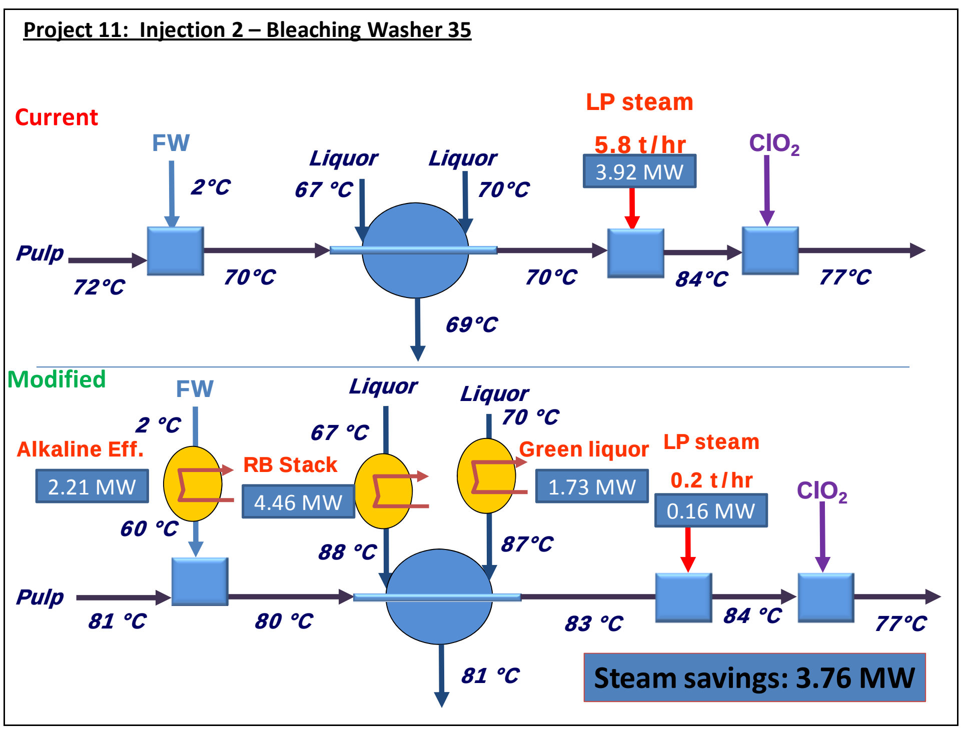


Figure 6-25: Project 11 – Injection 2 – Washer 35

Project 12 – Injection 3 – Washer 45:

In the current configuration, Fresh water is being injected into the pulp line resulting in a non isothermal mixing point of with a temperature drop of 1 °C. Washing liquors area injected into the washer at a temperature between 64-70 °C. Finally, steam is being injected in the bleaching pulp line to heat the pulp to 79 °C. The proposed design suggests the heating of the fresh water, washing liquor 1 and 2 to 60 °C, 88 °C and 87 °C respectively using acidic effluent and recovery boiler stack gases. By doing so, the temperature of the pulp exiting the washer increases by 10 °C which will result in a steam saving of 3.83 MW. One thing to note is that the heat exchanger used to heat liquor to 70 °C is the same as the heat exchanger used in project 11 to heat liquor at 70 °C. In addition, the heat exchanger used to heat liquor at 64 °C is the same as the heat exchanger used in project 13 to heat liquor at 64 °C. The project schematic is presented in figure 6-26.

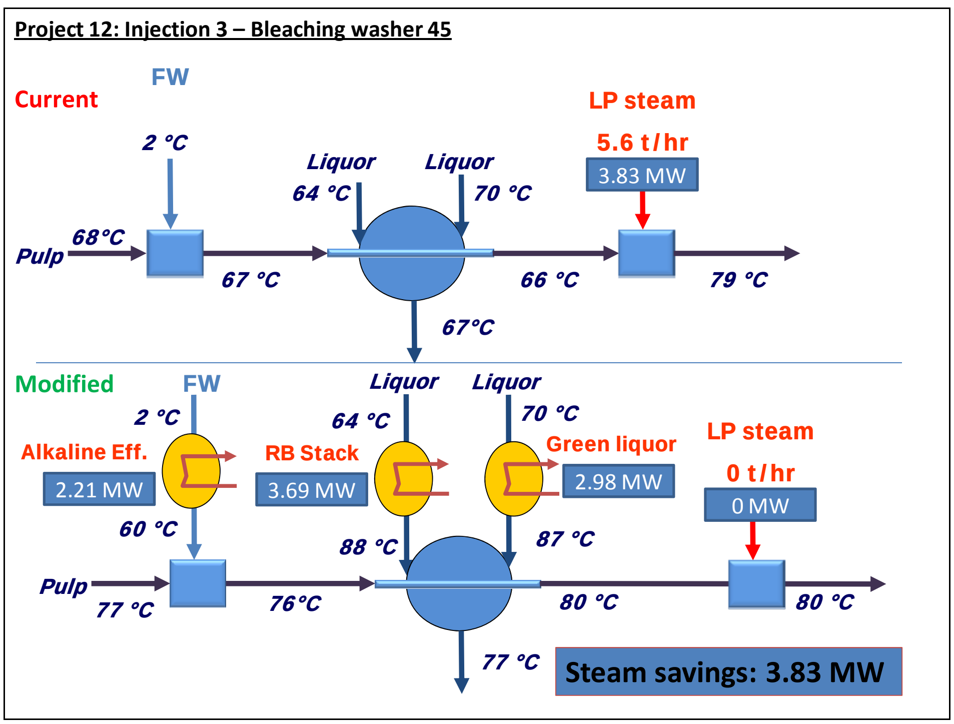


Figure 6-26: Project 12 – Injection 3 – Washer 45

Project 13 – Injection 4 – Washer 55:

In the current configuration, Washing liquor and white water are injected into the washer at a temperature between 57 - 64 °C. This results in a non isothermal mixing and a temperature drop in the pulp line of about 7 °C. Steam is being directly injected in the bleaching pulp line to heat the pulp to 79 °C. The proposed design suggests heating, washing liquor 1 and 2 to 88 °C and 82 °C respectively using recovery boiler stack gases. By doing so, the temperature of the pulp exiting the washer increases by 5 °C which will result in a steam saving of 2.92 MW. The pulp temperature increases 9 °C above target point; this will lead to further steam reduction in the pulp machine steam showers. One thing to note is that the heat exchanger used to heat white water at 57 °C is the same as the heat exchanger used in project 10 to heat white water. In addition, the heat exchanger used to heat liquor at 64 °C is the same as the heat exchanger used in project 12 to heat liquor at 64 °C. The project schematic is presented in figure 6-27.

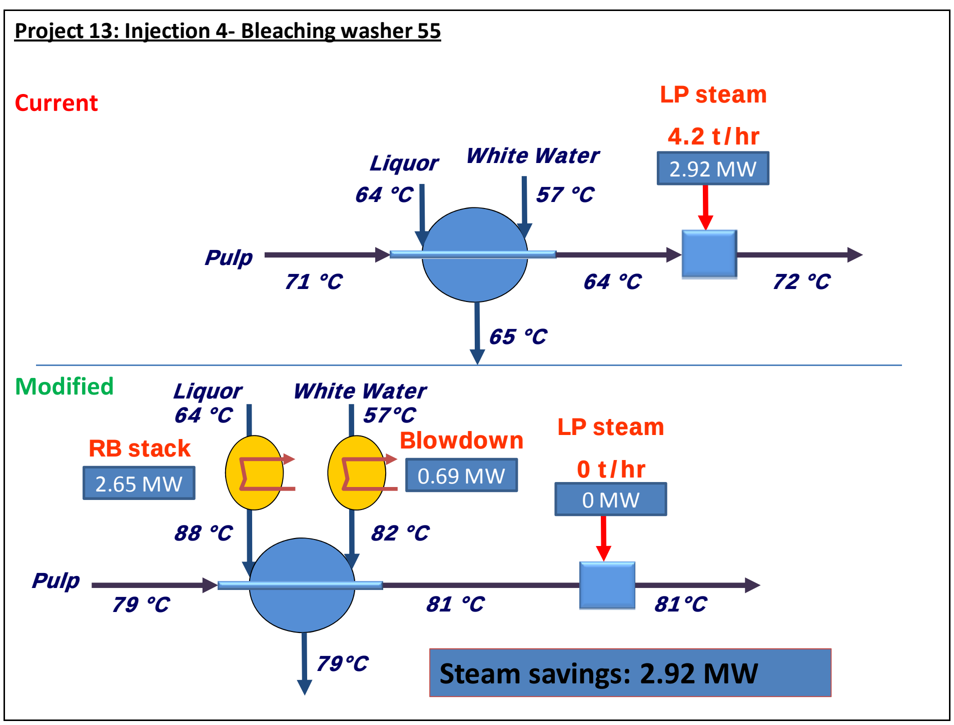


Figure 6-27: Project 13 – Injection 4 – Washer 55

Project 14 – Boiler Air Heater:

In the current configuration boiler heater uses low pressure (LP) steam to heat Air entering into the recovery boiler in steam plant. The temperature of the air increases from 45 °C to the target temperature of 80 °C. The proposed design suggests the replacement of LP steam by alkaline effluent below the pinch and recovery boiler stack gas above the pinch. The steam savings incurred are 3.16 MW. The project schematic is presented in figure 6-28.

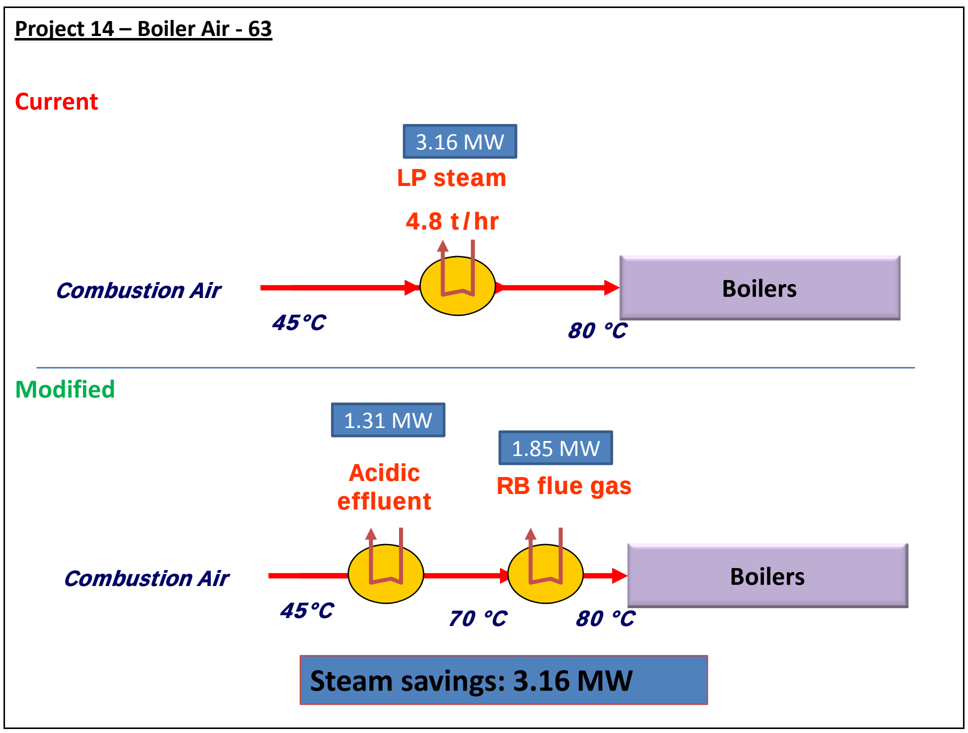


Figure 6-28: Project 14 – Boiler Air Heater

6.4.3.1 Summary of potential energy savings – Grassroot high savings

The theoretical targets obtained by the thermal pinch curves of the grassroot constraint levels presented higher energy savings than the retrofit constraint level. This is evident by the total savings based on the potential projects as well which equal to 32.76 MW or 17 % is of total steam consumed. Similar to the retrofit constraint level, the most promising projects in terms of energy savings are replacing the bleach heater and heating the makeup water. The main difference in potential energy targets is due to the increased savings in the bleaching steam injection projects. This is mainly because hot streams were used in more efficient design thus releasing more useful energy that could be used in the projects. The summary of the energy saving projects are presented in table 6-7.

Table 6-7: Summary of potential energy savings – Grassroot A

#	Project name	Steam Savings (MW)	Steam Savings (%)
1-8	**Bleach heater**	**7.37**	**3.9**
1-8	**Brown heater**	**3.24**	**1.7**
9	**Deaerator (make up water)**	**6.37**	**3.2**
10	**Injection 1 (Washer 15)**	**2.11**	**1.1**
11	**Injection 2 (Washer 35)**	**3.76**	**1.9**
12	**Injection 3 (Washer 45)**	**3.83**	**1.9**
13	**Injection 4 (Washer 55)**	**2.92**	**1.5**
14	**Boiler Air Heater**	**3.16**	**1.7**
Total		**32.76**	**17**

6.4.4 Summary of heat exchanger networks –Line A

Table 6-8 represents the results from the three heat exchanger networks at three different constraint levels; retrofit low, retrofit medium, and grassroot high. The energy savings increase from 11%, 15 %, and 17% as constraints becomes less rigid and more flexible. In addition, the number of heat exchangers and total extra area required for heat transfer increases as well. In the retrofit low constraint level, 3 existing heat exchangers were targeted and replaced with 8 new heat exchangers thus resulting in a total of 21 heat exchangers with an extra area of 2905 m^2. In the retrofit medium constrain level, 7 exiting heat exchangers were targeted. Green liquor cooler and bleach heaters were salvaged while the other five were replaced with 16 new heat exchangers resulting in a total of 29 with an extra area of 4202 m^2. Heat exchangers were salvaged based on the areas; if the area is within 5 % of the new heat exchanger then it can be upgraded and used. In the grassroot high constrain level, 9 exiting heat exchangers were targeted. Surface condenser 2 in evaporators was salvaged while the other8 were replaced with 21 new heat exchangers resulting in a total of 30 with an extra area of 9165 m^2. Economic analysis will be required to further analyze the impact of each constraint level and is presented in the last section.

Table 6-8: Summary of savings at different constraint levels – Line A

Data	Current	R-Low A	R-Med A	G-High A
Steam Savings (MW)	**-**	**21.5**	**28.7**	**32.8**
Steam Savings (%)	**-**	**11**	**15**	**17**
Total # of heat exchangers	**16**	**21**	**29**	**30**
exchangers that could be salvaged	**-**	**0**	**2**	**1**
heat exchangers that could be salvaged		**3**	**5**	**8**
# new heat exchangers	**-**	**8**	**16**	**21**
Extra Area Required (m^2)	**-**	**2905**	**4202**	**9165**

6.4.5 Line B – Retrofit – Medium savings

In a similar fashion and context to line A, retrofit – medium savings constraint level is applied to line B. Similar adjustments are made to the existing heat exchanger network by targeting the cross pinch heat exchangers. The energy liberated is then used to eliminate or reduce indirect steam consumption in heaters. In addition, Direct steam injection is reduced by eliminating non isothermal mixing points that are located in the deaerator, white water tank, dilution conveyer hot and warm water tank, and bleaching steam injection points 1-4. To be more specific, heating water or liquor streams with liberated hot process streams, effluents and stack gases to eliminate non isothermal mixing, low pressure steam consumption is reduced .by other. The proposed potential energy saving projects are presented below:

Project 1 – Boiler Air Heater B:

In the current configuration boiler heater uses low pressure (LP) steam to heat Air entering into the recovery boiler in steam plant. The temperature of the air increases from 45 °C to the target temperature of 80 °C. The proposed design suggests the replacement of LP steam by acidic effluent below the pinch and recovery boiler stack gas above the pinch. The steam savings incurred are 3.16 MW. The project schematic is presented in figure 6-29.

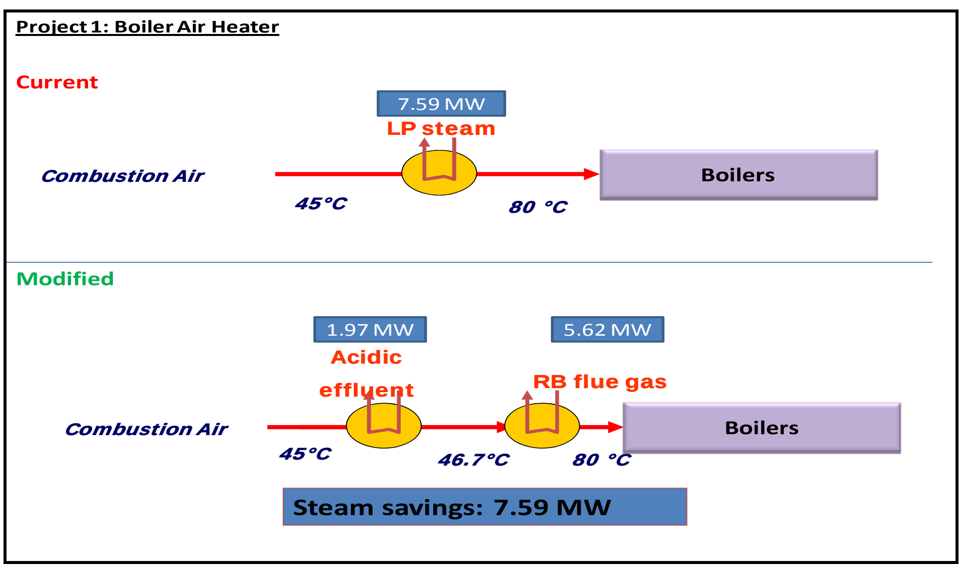


Figure 6-29: Project 1 – Boiler Air Heater B

Project 2 – Make up water (Deaerator) B :

In the current configuration, boiler make up water at 31 °C is mixed with condensate at 101 °C in condensate collection tank resulting in a non isothermal mixing point. Water at 65 °C is sent from the tank into the deaerator to be heated with direct steam. The proposed design focuses on the elimination of the non isothermal mixing in the tank thus reducing the steam injection in the deaerator. This can be achieved by heating the makeup water to 88 °C using acidic effluent below the pinch and recovery boiler stack gases above the pinch. The steam injected into the deaerator is reduced and a saving of 14.25 MW is obtained. The project schematic is presented in figure 6-30.

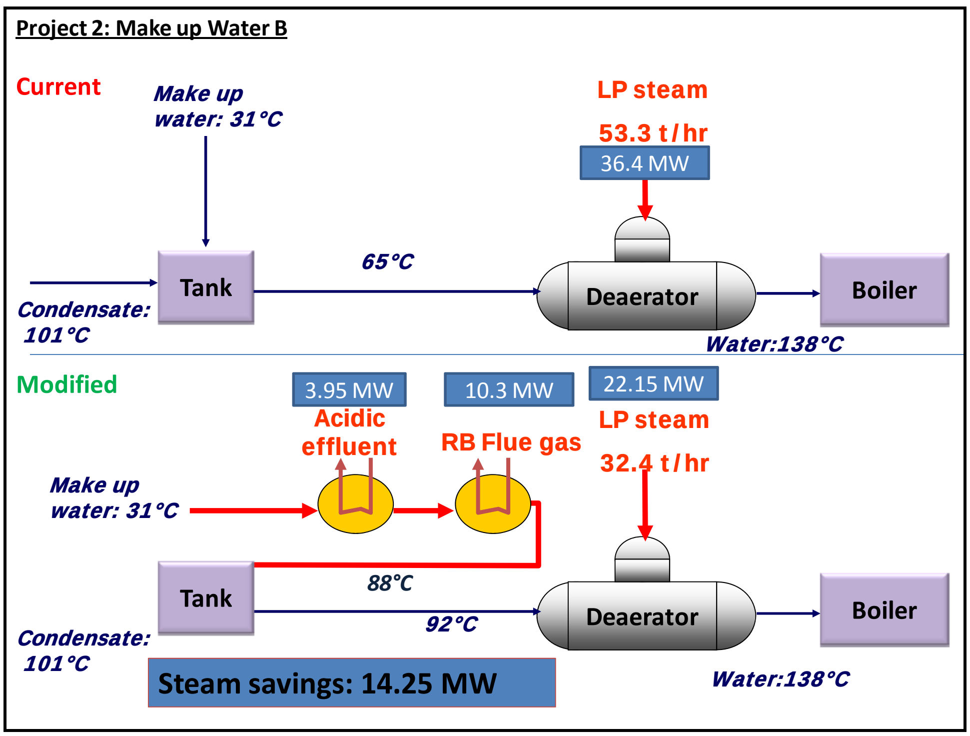


Figure 6-30: Project 2 – Make up water (Deaerator) B

Project 3 – Non isothermal mixing in dilution conveyer:

In the current heat exchanger network R8 ClO_2 chemical is cooled down in an indirect condenser using fresh water. The produced water at 50 °C is sent to the warm water tank. A pinch violation of 2.50 MW is noticed due to this heat transfer. In the dilution conveyer in brownstock washing, pulp is diluted with warm water and hot water before being sent to the bleaching. The modified design suggests using acidic effluent to heat warm water below the pinch and thus releasing the R8 ClO_2 chemical. The ClO_2 chemical is then used to heat warm water to 76 °C before it enters the dilution conveyer. The steam savings due to this project are 0.8 MW. The project schematic is presented in figure 6-31.

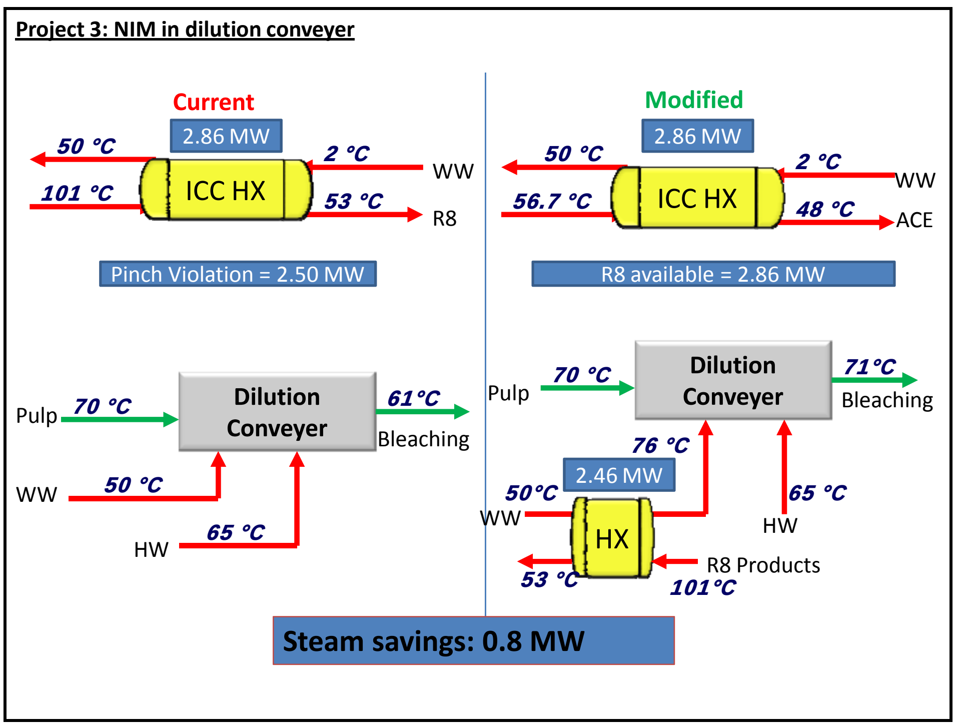


Figure 6-31: Project 3 – Non isothermal mixing in dilution conveyer

Project 4 – Non isothermal mixing in white water tank:

In the current configuration, LP steam is being injected into the white water tank to achieve a target temperature in the tank of 77 °C. LP steam is used to compensate for the loss of temperature due to the addition of makeup fresh water at 2 °C. A non isothermal mixing occurs due to this heating strategy. The modified scenario proposes a project where by fresh water is heated below the pinch using evaporator's effluents while above the pinch exhaust air from the machine is used as a heating source. If the project is implemented, savings of 2.93 MW of LP steam could be achieved. The schematic of the project is presented in figure 6-32.

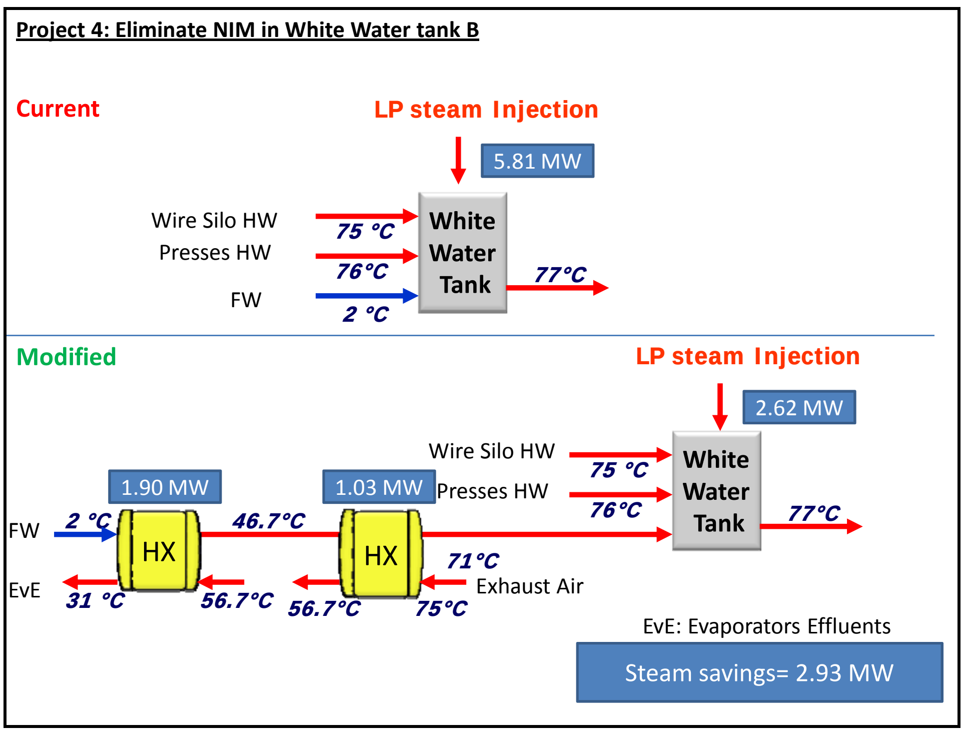


Figure 6-32: Project 4 – Non isothermal mixing in white water tank

Project 5 – Bleach heater and Direct condenser:

The bleach heater and direct condenser are currently being used to heat hot water and warm water respectively. The bleach heater uses indirect steam heater to bring the water temperature from hot water tank to bleaching to 65 °C. On the other hand, the direct condenser uses direct steam injection to heat water from warm water tank to hot water tank to a temperature of 84 °C. Flashed steam condenser uses flashed steam from flash tanks in the digester department to heat warm water to 55 °C before sending to the hot water tank. This practice results in a non isothermal mixing in the hot water tank and a degradation of good quality energy due to the use of direct steam injection. In the modified configuration, warm water from the warm water tank is heated in two heat exchangers to a temperature of 61.5 and 65.8 respectively. The hot sources used are acidic effluents and alkaline effluents. By implementing this project, Non isothermal mixing in the hot water tank is eliminated, steam direct and indirect injection is eliminated, and energy from flashed steam of 2.86 MW is released to be used in project 6. The steam savings due to this modified practice is 7.22 MW. The project schematic is presented in figure 6-33.

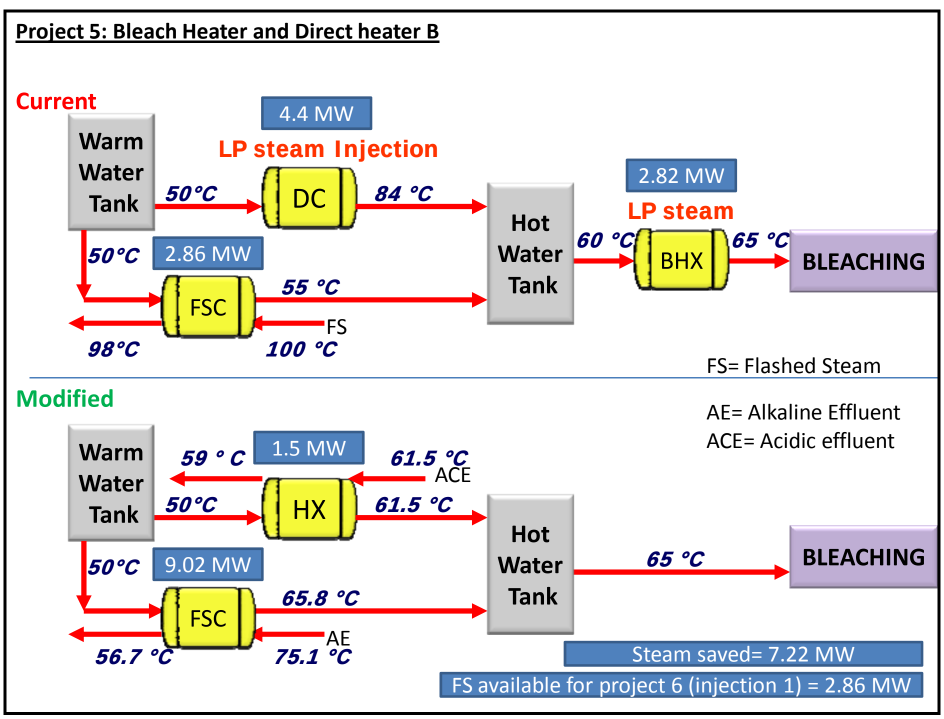


Figure 6-33: Project 5 – Bleach heater and Direct condenser

Pre Project – Elimination of violations in cold blow cooler and green liquor cooler:

The mill in its current configuration uses fresh water to cool down cold blow liquor to a temperature of 70 °C. This practice results in a huge pinch violation equals to 10.38 MW. In addition green liquor is being cooled with fresh water to a temperature of 94 °C causing a pinch violation of 2.92 MW. The idea behind this project is to release the energy in both heat exchangers to be used in a more efficient manner. In cold blow cooler case, fresh water is being heated by alkaline effluent below the pinch and cold blow liquor above the pinch thus releasing 10.38 MW of energy to be used in bleaching steam injection projects. In the case of green liquor cooler, fresh water is being heated with alkaline effluent below the pinch and green liquor above the pinch. The energy released in green liquor is 2.92 MW and it is to be used in bleaching steam injection projects. The project schematic is presented in figure 6-34.

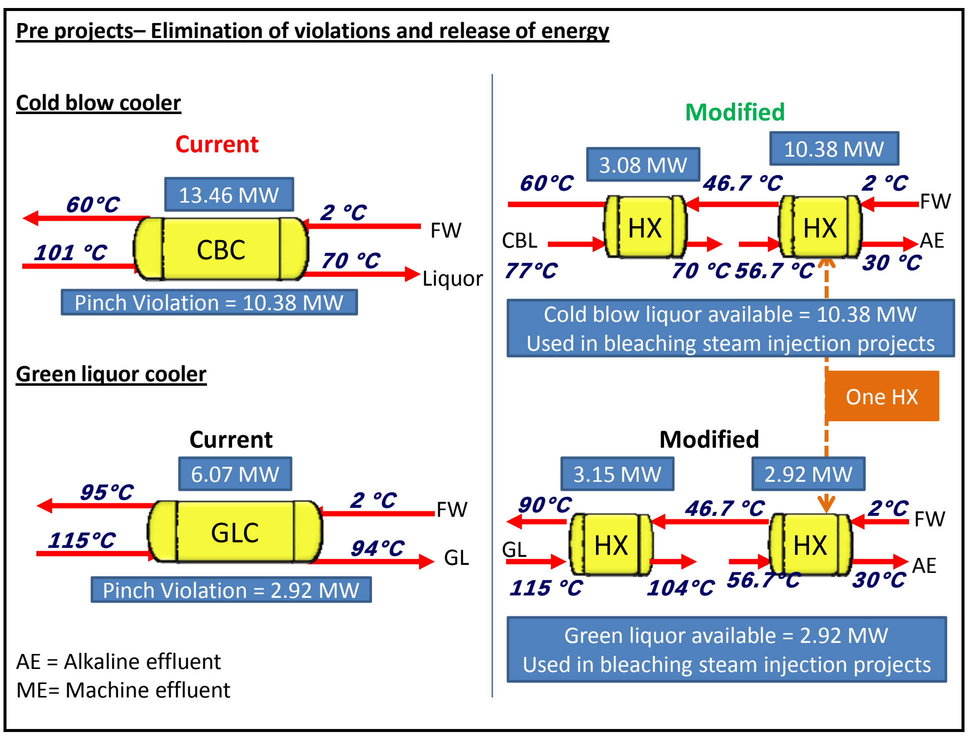


Figure 6-34: Pre Project – Elimination of violations in cold blow cooler and green liquor cooler

Project 6 – Injection 1 – Washer 15 B:

In the current configuration, Washing liquor and hot water are injected into the washer at a temperature between 65-70 °C. Steam is being directly injected in the bleaching pulp line to heat the pulp to 84 °C resulting in a non isothermal mixing point. The proposed design suggests heating, washing liquor and hot water to 84 °C and 84 °C respectively using cold blow liquor and flashed steam from digester department. By doing so, the temperature of the pulp exiting the washer increases by 11 °C which will result in a steam saving of 3.47 MW. One thing to note is that the heat exchanger used to heat liquor at 70 °C is the same as the heat exchanger used in project 7 to heat the washing liquor. The project schematic is presented in figure 6-35.

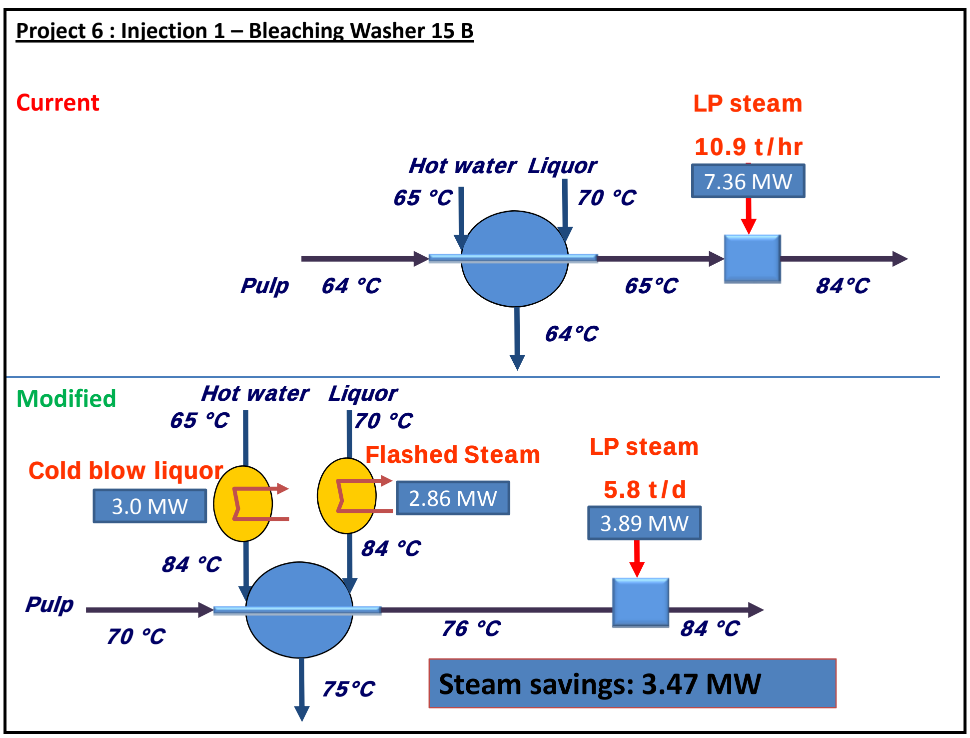


Figure 6-35: Project 6 – Injection 1 – Washer 15 B

Project 7 – Injection2 – Washer 35 B:

In the current configuration, both washing liquor streams are being injected into the washer at a temperature between 70-72 °C. Steam is being directly injected in the bleaching pulp line to heat the pulp to 85 °C resulting in a non isothermal mixing point. The proposed design suggests heating, washing liquor 1 and washing liquor 2 to 83 °C and 84 °C respectively using cold blow liquor and green liquor. By doing so, the temperature of the pulp exiting the washer increases by 7 °C which will result in a steam saving of 2.17 MW. One thing to note is that the heat exchanger used to heat liquor at 70 °C is the same as the heat exchanger used in project 6 to heat the washing liquor. In addition, the heat exchanger used to heat liquor at 72 °C is the same as the heat exchanger used in project 8 to heat liquor at 72 °C. The project schematic is presented in figure 6-36.

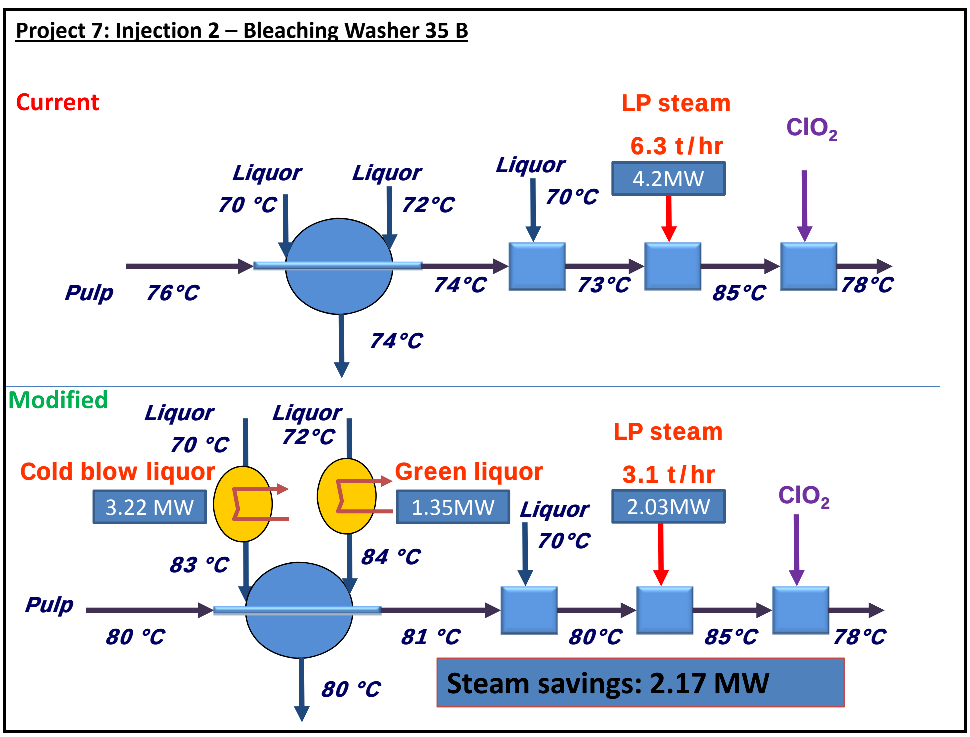


Figure 6-36: Project 7 – Injection2 – Washer 35 B

Project 8 – Injection 3 – Washer 45 B:

In the current configuration, both washing liquor streams are being injected into the washer at a temperature between 69-72 °C. Steam is being directly injected in the bleaching pulp line to heat the pulp to 82 °C resulting in a non isothermal mixing point. The proposed design suggests heating, washing liquor 1 and washing liquor 2 to 84 °C and 79 °C respectively using green liquor and cold blow liquor. By doing so, the temperature of the pulp exiting the washer increases by 3 °C which will result in a steam saving of 1.88 MW. One thing to note is that the heat exchanger used to heat liquor at 72 °C is the same as the heat exchanger used in project 7 to heat the washing liquor. In addition, the heat exchanger used to heat liquor at 69 °C is the same as the heat exchanger used in project 9 to heat liquor at 69 °C. The project schematic is presented in figure 6-37.

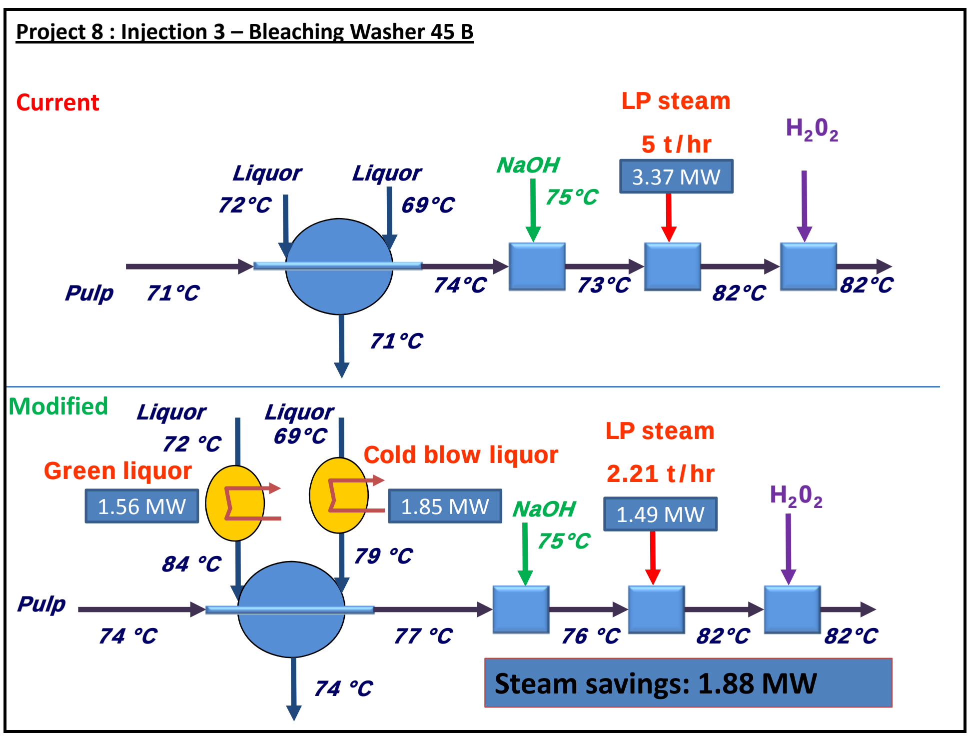


Figure 6-37: Project 8 – Injection 3 – Washer 45 B

Project 9 – Injection 4 – Washer 55 B:

In the current configuration, both washing liquor stream and hot water are being injected into the washer at a temperature between 65-69 °C. Steam is being directly injected in the bleaching pulp line to heat the pulp to 78 °C resulting in a non isothermal mixing point. The proposed design suggests heating washing liquor 1 and hot water to 79 °C using cold blow liquor and blowdown. By doing so, the temperature of the pulp exiting the washer increases by 7 °C which will result in a steam saving of 2.54 MW. One thing to note is that the heat exchanger used to heat liquor at 69 °C is the same as the heat exchanger used in project 8 to heat the washing liquor. The project schematic is presented in figure 6-38.

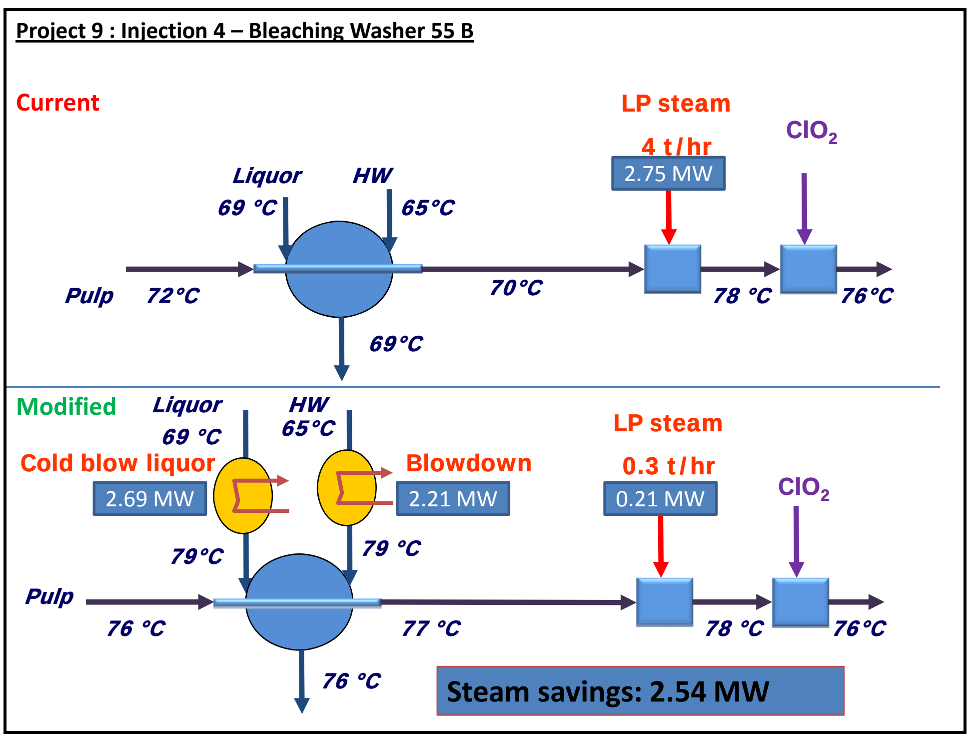


Figure 6-38: Project 9 – Injection 4 – Washer 55 B

6.4.5.1 Summary of potential energy saving projects – Retrofit medium B

The most promising projects with respect to energy savings are internal heat recovery in makeup boiler water and boiler air heater. These two projects add up to a staggering value of 8.1 % of total savings. The other projects have savings ranging between 0.3% and 2 %. The total energy saved based on the compilation of the projects is 42.87 MW or 16% of total steam consumption. The summary of the energy saving projects are presented in table 6-9.

Table 6-9: Summary of potential energy saving projects – Retrofit medium B

#	Project name	Steam Savings (MW)	Steam Savings (%)
1	**Boiler Air Heater**	**7.59**	**2.8**
2	**Deaerator (make up water)**	**14.25**	**5.3**
3	**NIM in Washing Dilution conveyer**	**0.81**	**0.3**
4	**NIM in White Water tank**	**2.93**	**1.1**
5	**Bleach heater**	**2.82**	**1.1**
	Direct Condenser	**4.40**	**1.6**
6	**Injection 1 (washer 15)**	**3.47**	**1.3**
7	**Injection 2 (washer 35)**	**2.17**	**0.8**
8	**Injection 3 (washer 45)**	**1.88**	**0.7**
9	**Injection 4 (washer 55)**	**2.54**	**0.9**
Total		**42.87**	**16**

6.4.6 Summary of heat exchanger network - Line B Retrofit medium

At the retrofit medium constraint level, a heat exchanger network was developed with 42.9 MW or 16 % of energy savings based on the total consumption. In the medium savings heat exchanger network, 6 exiting heat exchangers which had pinch and crisscross violations were targeted. Green liquor cooler and cold blow cooler were salvaged while the other heat exchangers were replaced with 16 new ones resulting in a total of 31 with an extra area of 12390 m^2. Heat exchangers were salvaged based on the areas; if the area is within 5 % of the new heat exchanger then it can be upgraded and used. It is apparent that the area to saving ratio in line B is higher than line A; this is due to the fact that different streams with varying heat transfer coefficients are used in the heat exchangers which will affect the size of each heat exchanger significantly. Economic analysis will be presented in the next section to further analyze the impact of the new heat exchanger on the mill profitability. Table 6-10 contains the results from the retrofit medium savings constraint level for line B.

Table 6-10: Summary of heat exchanger network - Retrofit medium B

Data	Current	Retrofit -Med B
Steam Savings (MW)		**42.9**
Steam Savings (%)		**16**
Total # of heat exchangers	**17**	**31**
heat exchangers that could be salvaged		**2**
Number of heat exchangers that could not salvaged		**4**
Number new heat exchangers		**16**
Extra Area Required (m^2)		**12390**

6.5 Economic Analysis:

The economic evaluation of the proposed heat exchanger networks for line A and line B is presented in this section. The first parameter that is calculated is the capital cost of the new heat exchangers. The cost of each heat exchanger is calculated based on the following generic formula[26]:

$$Capital\ Cost\ (\$) = C_B * F_D * F_P * F_M \qquad (1)$$

Whereby:

C_B is base cost for a carbon steel floating head heat exchanger and is calculated based on the following formula:

$$C_B = e^{[8.202+0.01506(\ln Area) + 0.06811\,(ln\,Area)^2]} \quad (2)$$

F_D is the exchanger type cost factor when switching from a floating head to a fixed head and is calculated based on the following formula:

$$F_D = e^{[-0.9003+0.0906\,(ln\,Area)]} \qquad (3)$$

F_P is the design pressure factor to handle pressures up to 4000 KPa and is calculated based on the following formula:

$$F_P = 1.4272 + 0.12088\,[(ln\,Area)]\ (4)$$

F_M is the material cost factor for stainless steel and titanium heat exchangers

$F_M = 1.4144 + 0.23296\,(ln\,Area)$ for Stainless steel 316 (5)

$F_M = 2.5617 + 0.42913\,(ln\,Area)$ for Titanium (6)

Heat exchangers that use recovery boiler/power boiler stack gas or lime kiln stack gas are made up of titanium while all the other heat exchangers are made up of stainless steel 316.

The total capital cost of the heat exchanger network is evaluated by adding the cost of each heat exchangers plus the piping and installation which is 31% of the total capital cost[26]. Finally, the cost is actualized to $ 2011 using a simple cost index to adjust it from 1989 to 2011[27]. The data and the formula are found in appendix 3-2.

The steam savings from the mill could be transformed in three different scenarios to obtain operating cost savings. These scenarios area:

1- Reduction of fuel consumption:

This scenario is achieved by reducing the high pressure steam production from the power boiler. This will result in lower fuel consumption which are primarily Hog and natural gas. One thing to note is the reduction of steam production will affect the total electricity produced from the turbine. A s a result, the net gain due to the reduction of fuel should be calculated by factoring out the effect of purchasing more electricity from the grid. In addition, the amount of stack gases produced from the power boiler will be reduced as well. The produced steam from power boiler uses natural gas and hog as fuel. 95% of the produced steam uses hog while the other 5% uses natural gas.

2- Increase electricity production

The idea in this scenario is to use the extra low pressure steam in a condensing turbine to increase the total electricity production and reduce the dependence on the electricity from the grid. The capital cost in this scenario will increase drastically due to the purchase and installation of a new condensing turbine. If the condensing turbine already exists in the mill, the situation would have been much more promising.

3- Biorefinery option

The third scenario involves the use of the steam in a future project which is the implementation of biorefinery unit in the mill. This will help to keep the operating cost of the mill constant while producing a high value product to increase the total revenues. This scenario will not be evaluated.

The capital cost, operating cost savings, and simple payback period is presented in the next section for each of the heat exchanger networks built for line A and line B. More information regarding the detailed calculation is presented in appendix 3-3 and 3-4.

6.5.1 Heat exchanger networks - Line A

In either steam saving scenario, it is evident that as the level of constraints changes from retrofit to grassroot, the capital cost and the operating cost savings increases. The increase in operating savings is evident between retrofit- low and retrofit medium than retrofit medium and grassroot.

By considering the scenario where steam production is reduced, all three constraint levels show a promising side where by the simple payback period increases from 1.7 years to 3 years. In the second scenario where electricity production is increased, the simple payback period increases from 5.6 years to 6.7 years making this scenario less appealing. Therefore the option of reducing steam production in any of the constraint levels requires three years or less to repay the initial investment. The list of heat exchangers and detailed economical analysis for line A projects are available in appendix 3-3. The information is summarized in figure 6-39.

Figure 6-39: Steam savings economical scenarios Line A

6.5.2 Heat exchanger network - Line B

The retrofit medium savings heat exchanger network was only built for line B due to the initial findings from the composite curves. The results in this line is consistent to the results shown in line A whereby the scenario of reducing steam production is more economically appealing than the scenario of increasing electricity production. The payback period for reducing the steam production scenario is 3.6 years versus 7.4 years for the scenario where by electricity production is increased. The increase in capital cost due to the purchase of a condensing turbine is not justified by the small increase in operating cost savings. Therefore the scenario where steam production is produced is more profitable in the short run. The list of heat exchangers and detailed economical analysis for line B projects are available in appendix 3-4. The information is summarized in figure 6-40.

Figure 6-40: Steam savings economical scenarios

CHAPTRE 7 CONCLUSION AND RECOMMENDATIONS

7.1 Conclusions

Based on the analysis, the hypotheses that were set at the beginning of the projects were not refuted. In terms of the model, the discrepancy in total steam and water production and consumption between model and data is less than 5%. The mill has been characterized and and points of inefficiencies have been identified. It was proven that grassroot approach has an effect on the total energy savings in line A. The potential energy saving projects and the heat exchanger networks in line A and line B were viable from an economic point of view.

Theoretical savings based on the composite curves for line A were 22% in grassroot and 20% in retrofit. Based on the different heat exchanger network designs, it was possible to achieve 17% savings in grassroot and a maximum of 15% in retrofit. For line B, theoretical savings based on the composite curves were 24% and the maximum potential savings based on the heat exchanger network design was 16%.

The economics for the heat exchanger networks of line A, show that for the retrofit case, a simple payback period of 2.1 years is achievable while for the grassroot case a simple payback period of 3 years is achievable. This is the case when the chosen scenario is reducing the steam production and fuel consumption. Therefore, one can say that it is economically viable to design either in grassroot or retrofit constraint level for line A. For line B, the retrofit heat exchanger network could be built with a simple payback period of 3.6 years if steam production is the chosen scenario. Increasing the steam production to produce more electricity was not an economically feasible scenario for both line A and line B.

Constraint analysis was performed on the overall mill based on a systematic and documented approach. A set of guidelines have been developed in order to customize the constraint analysis process to any pulp and paper mill. The guidelines will enable future users to locate energy saving projects systematically and achieve realistic and potential energy targets in short period of time.

7.2 Recommendations

Further work should be done in order to screen the potential projects and obtain a list of practical projects. This will require in depth revision of the projects with the mill personnel to determine which of the proposed projects could be applied to the mill.

Combined water and energy analysis should be done to determine the potential water savings in the process. The proposed water project will change the energy demand of the mill and therefore could affect the already proposed energy projects. Interaction analysis between the water and energy projects should be done to determine if more savings could be achieved due to the implementation of water projects.

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ANNEXE 1 – Simulation model, Validation, and Characterization

1-1 Steam network tables

1-2 Injected steam

1-3 Water network tables

1-1 Steam network tables

Steam Network Table – Line A						
Stream	Description	m [t/h]	E_{sp} [kJ/kg]	Q [kJ/s]	P (kPa)	t [°C]
V1a	HP produced from BB2	0.0	20895	0	4300	4000
V2a	HP produced from RB1	193.8	3205	172500	4300	400
V3a	LP consumed in Deaerator	23.5	2792	14538	450	170
V4a	HP entering through MP PRV	0.0	0	0	0	0.0
V5a	HP entering through LP PRV	0.0	3205	0	4300	4000
V6a	LP produced from turbine	190.9	2848	151027	450	195
V7a	MP produced from turbine	77.6	2982	64305	1150	271
V8a	LP consumed in Digester	13.8	2792	10687	450	170
V9a	MP Consumed in Digester	10.3	2822	8080	1150	202
V10a	MP consumed in Washing	5.6	2822	4401	1150	202
V11a	LP consumed in Pulp Machine	31.5	2792	24401	450	170
V12a	MP consumed in Pulp Machine	50.1	2822	39262	1150	202
V13a	LP consumed in Bleaching	21.0	2792	16253	450	170
V14a	LP consumed in Evaporators	61.9	2792	47258	450	170
V15a	LP consumed in Water treatment	17.3	2792	13413	450	170
V16a	LP consumed in Air HX	4.8	0	3721	1150	202
V17a	LP consumed in recaust	0.0	2792	0	450	170
V18a	MP consumed in evaporators	1.8	2822	1411	1150	202
v19a	Extra MP at the end of the line	0.2	2822	124	1150	202
v20a	Extra LP at the end of the line	0.8	2792	585	450	170
Vab1	HP from line B	74.8	3205	66589	4300	400
Vab2	MP from line B	15.4	2822	12085	1150	202
Vab3	LP sent to line A	21.2	2792	16451	450	170
C1a	CD produced in digester	8.8	462	1132		109
C2a	CD Produced in Washing	0.0	0	0		0.0
C3a	CD produced in Pulp Machine	42.7	462	5487		110
C4a	CD from evaporators	61.9	418	7192		99.6
C5a	CD produced in water prod	15.7	420	1836		100
C6a	CD produced from steam plant	0.0	420	559		100
C7a	Total CD return	134.0	435	16207		103
De1a	CD sent to PRV's	10.3	580	1666		137
De2a	CD sent to Deaerator	206.3	328	18798		78.4
De3a	CD sent to recovery boilers	229.8	580	37057		137
De4a	CD sent to RB1	219.3	580	35358		137
De5a	CD sent to BB2	0.0	580	0		137

FW1a	FW entering CD tank	72.2	129	2588		31.0

Steam Network Table – Line B

Stream	Description	m [t/h]	E_{sp} [kJ/kg]	Q [kJ/s]	P (kPa)	t [°C]
V1b	HP produced from BB4	127.9	3205	113887	4300	400
V2b	HP Produced from RB5	248.6	3205	221357	4299	400
V3b	LP consumed in Deaerator	52.0	2792	40301	450	170
V4b	HP consumed in LP PRV	7.8	3205	6945	4300	400
V5b	HP consumed in MP PRV	1.8	3205	1595	4300	400
V6b	LP produced from turbine	230.1	2833	181039	450	188
V7b	MP produced from turbine	62.1	2974	51296	1150	268
V8b	LP consumed in steaming vessel	27.2	2792	21134	450	170
V9b	MP consumed in liquor HX	29.0	2822	22759	1150	202
V10b	LP consumed in bleaching	26.1	2792	20263	450	170
V11b	LP consumed in chem. prep	6.3	2792	4848	450	170
V13b	LP consumed in evaporators	77.7	2792	59120	450	170
V14b	LP consumed in Air HX in S.P	5.9	2792	4598	450	170
	LP consumed in condensate stripper	17.0	2792	13186	450	170
	LP consumed in space heating	19.0	2792	14738	450	170
V15b	MP consumed in unknown S.P	5.0	2822	3919	1150	202
V16b	LP consumed in Water Production	8.5	2792	6571	450	170
V17b	LP consumed in PM	25.0	2792	19392	450	170
V18b	MP consumed in PM	47.8	2822	37479	1150	202
V19b	LP at the end of the line	0.0	2792	24	450	170
V20b	MP at the end of the line	0.1	2792	105	450	170
v21b	MP consumed in evaporators	1.8	2822	1088	450	202
C1b	CD produced in the digester	24.8	462	3181		109
C3b	CD produced in Evaporators	30.8	420	3593		100
C4b	CD produced in Ea from Cond 2	46.9	424	5525		101
C5b	CD produced from space heating	9.6	420	1119		100
C6b	CD produced P.M MP steam	47.0	462	6028		109
C7b	CD produced from Reboiler in CP	5.7	420	659		100
C8b	CD from air HX in SP	5.9	420	691		100
C10b	CD from bleach HX in water prod.	4.3	420	497		100
C9b	Total CD back	174.9	428	20796		101
De1b	CD to PRVS	10.2	580	1641		137
De2b	CD to deaerator	370.3	270	27793		64.7
De3b	CD to boilers	422.3	580	68094		137
De4b	CD to RB	281.4	580	45372		137

De5b	CD TO BB	130.7	580	21081		137
FW1b	FW consumed in CD Tank	195.6	8	451		2.0

1-2 Injected steam tables

Injected steam – Line A						
Stream	Description	m [t/h]	E_{sp} [kJ/kg]	Q [kJ/s]	P (kPa)	t [°C]
V8a	LP into steaming vessel	13.8	2792	10686	450	170
v21a	MP injected in washing	5.6	2822	4400	1150	202
V13a	LP injected directly in bleaching	21.0	2792	16253.2	450	170
v22a	LP injected directly in P.M	31.5	2792	24401	450	170
V17a	LP injected into recaustizing	0.0	2792	0.0	450	170
v23a	LP from direct heat exchanger in WT	0.0	2792	0.0	450	170
V3a	LP injected into deaerator	23.5	2792	14538.1	450	170
v18a	MP consumed in evaporators	1.8	2822	1410	1150	202

Injected steam – Line B						
Stream	Description	m [t/h]	E_{sp} [kJ/kg]	Q [kJ/s]	P (kPa)	t [°C]
V8b	LP into steaming vessel	27.2	2792	21134	450	170
V10b	LP injected directly in bleaching	26.1	2792	20262	450	170
V21b	LP injected directly in P.M	25.0	2792	19391	450	170
V22b	LP Steam injected in C.P	0.6	2792	462	450	170
V14b	LP injected in condensate stripper	17.0	2792	13186	450	170
V15b	LP injected directly in recausticizing	5.0	2792	3919	450	170
V16b	LP steam in direct condenser in water production	4.2	2792	6570	450	170
V3b	LP steam injected into deaerator	52.0	2792	40300	450	170
	LP consumed in space heating	9.4	2792	7295	450	170

1-3 Water network tables

Water network tables A						
Stream	Description	E_{sp} [kJ/kg]	Q [kJ/s]	t [°C]	Dissolved Solids (%)	water (m3/hr)
Digester						
w1a	WW Heated in cold blow cooler and sent to HW tank	334.9	9636.3	80.0		106.6
w2a	WW heated in steam condenser and sent to HW tank	292.6	34394.7	70.0		432.7
w20a	water in pulp entering digester	0.2	3.7	0.1	4.1	12.9
Washing						
w3a	WW consumed in washing	237.7		56.9		59.1
w14a	water in NaOH in washing	119.6	276.9	30.0	6.0	7.8
Bleaching						
w4a	Water from stripper cond in s.p B	276.6		66.2		88.2
w5a	FW consumed in bleaching	8.3	172.9	2.0		75.0
w6a	WW consumed in bleaching	237.7	3902.3	56.9		60.0
w7a	HW consumed in bleaching	334.9	66528.0	80.0		735.9
w15ap	peroxide	154.4	192.83	40	10.49	4.02
w15as	sodiumhydorxide	291.8	1217.37	75	9.95	13.52
w15ac	chlorine dixide	61.9	1264.55	15	0.80	72.86
w15a- Total	water in bleaching chemicals		2674.7			90.4
Pulp Machine						
w8a	HW consumed in pulp machine	334.9	2325.6	80.0		25.7
w23a	FW consumed in Pulp Machine	8.2	6.06E+02	2	0	262.8

	Overflow from wwchest B to A					35
Stream	Description	E_{sp} [kJ/kg]	Q [kJ/s]	t [°C]	Dissolved Solids (%)	water (m3/hr)
Evaporators						
w9a	FW heated in surf. Cond and sent to WW tank	280.0	38264.7	67.0		502.3
w22a	FW entering stripper condenser	8.2	4.15E+02	2	0	180
Water Prod						
w10a	FW heated in a direct condenser and sent to WW tank	8.3	141.1	2.0		61.2
w11a	HW recirculated through brown HX to HW tank	334.9	5814.0	80.0		64.3
w16a	HW water from B	250.5	7421.9	60.0		108.5
w17a	WW FROM B	207.1	12766.2	49.7		224.5
w21a	WW water from WW tank to HW tank	237.7	8593.3	56.9		132.2
Steam Plant						
w12a	FW consumed in Condensate collection tank	129.0	2588.4	31.0	0.0	72.6
w18a	water enetering with Bark in SP	377.2	0.0	90.0	0.0	0.0
Recausticizing						
w13a	Contaminated condensate from line B	4076.9	12850.7	65.2	0.0	173.3
w19a	water entering recaust with NaOH	52.7	4.5	20.0	50.0	0.2
w24a	FW after green liquor cooler	334.8		80		77.6
w25a	FW after dust vent scrubber exchanger	377.2		90		25.3
w26a	FW used in seal 1					22.7

	FW used in kiln cooling					17.8

Effluents						
Stream	Description	E_{sp} [kJ/kg]	Q [kJ/s]	t [°C]	Dissolved Solids (%)	water (m3/hr)
Digester						
E5a	NCG'S To turp recoery in digester	398.5	1130.5	95.0	0.0	10.2
Washing						
E1a	Effluent in washing leaving to sewers	317.0	433.3	77.4	1.2	4.8
E9a	over flow in washing	313.1	1109.9	76.5	1.9	12.2
	Effluent going to trash bin	254.4	228.5	76.4	1.9	2.3
Bleaching						
E6a	Acidic effluents in bleaching	302.3	57813.3	72.5	0.3	686.0
E7a	Alkali effluents in bleaching	289.5	32717.5	69.5	0.5	404.6
E10a	Overflow in contaminated condensate tank	296.8	2794.7	71.0	0.0	33.9
ba054	Overflow from wash press filtrate tank	343.8	2726.1	82.6	0.9	28.3
	Water sent to line B displacement press	296.8		71.0		141.7
Pulp Machine						
E2a	Effluent in pulp machine leaving to sewers	234.4	1365.1	56.6	0.0	20.8
E8a	Water leaving with pulp in PM	144.3	1342.7	90.0	0.5	3.3
e12al	Fresh water leaving vacuum pumps	8.30	276.62	2.00	0	120
E13a	water leaving with air from	723.7	3.19E+04	73.8	0.0	40.0

	dryer in PM					

Stream	Description	E_{sp} [kJ/kg]	Q [kJ/s]	t [°C]	Dissolved Solids (%)	water (m3/hr)
E14a	water leaving from noss screens in PM	229.4	530.9	55.4	0.0	8.2
Evaporators						
E3a	Effluent in evaporators leaving to sewers	91.1	5885	21.9	0.0	232.7
Steam Plant						
E17arb	Stack gases from recovery boiler	897.8	106586	252.83	0	93.56
E17abb	Stack Gases from bark boiler	443.1	0.00	105.49	0	0.00
Blowdown from recovery boiler	Blowdown from recovery boiler	1105	2664	253.7	0	10.91
Blow down from power boiler	Blow down from power boiler	1109	0	254.7		0.00
Recasuticizing						
E4a	Effluent in recaustification leaving to sewers	314.2	721.8	77.7	0.0	7.9
E15a	effluent from lime kiln cooling	49.4	333.9	11.9	0.0	24.4
E16aa	gases from lime kiln	1080.3	2318.7	250.00	0.03	7.67
E16ab	gases from causticizer	411.0	470.05	97.96	0.00	4.12
	effluent from seal water	8.30		2.00		17.81

List of heat exchangers – Line A

Heat Ex.	location	name
HX1	Digester	blow cooler
HX2	Digester	steam condenser
HX3	Evaporators	SURF COND 1,2
HX4	Water Production	direct Condenser
HX5	water production	Bleach heater
HX6	water production	brown heater

HX7	Recausticizing	Green Liquor Cooler
HX8	Recausticizing	Dust Vent Scrubber Exchanger

Water network tables B

Stream	Notes	E_{sp} [kJ/kg]	Q [kJ/s]	t [°C]	Dissolved Solids (%)	water (m3/hr)
Digester						
w1b	FW heated and sent to WW tank in cold blow cooler	251	13923	60.0		203.5
w3b	WW heated and sent to HW tank in flashed steam condenser	230	31203	55.0		496.5
w28b	Water entering with pulp	0	4	0.1	4.1	14.9
Washing						
w5b	HW consumed in washing	272	7174	65.0		97.0
w6b	WW consumed in washing	0	23098	50.0		101.1
w7b	Water from contaminated condensate tank from line A	297		71.0		141.7
Bleaching						
w8b	WW consumed in bleaching	0	3434	50.0		60.0
w9b	HW consumed in bleaching	272	28845	65.0		390.0
w22ba	peroxide	155	114	40.0	10.17	2.38
w22bb	NaOH	290	264	75.0	10.95	2.93
w22bc	chlorine dioxide	62	762	15.0	0.81	43.89
w22bd	sulphuric acid	42	86	10.0	0.00	7.50
w22b- Total	total water in bleaching chemicals		1227			56.7
Pulp Machine						
w10b	WW heated and consumed in P.M	272		65.0		91.4
w30b	Fresh water to seal pumps and white water chest	8	415	2.0	0.00	180.00
Evaporators						
w11b	FW heated in surf. cond sent to WW tank	196	42676	47.0		792.4
Water Prod						
w12b	HW sent to HW tank in line A	251	5799	60.0		84.8
w13b	WW sent to WW tank in line A	207	12766	49.7		224.5
w14b	WW heated in a direct HX and sent back to HW tank	353	9840	84.3		103.5

w24b	Water from HW tank to WW tank	251	1622	60.0	0.0	23.3
Stream	Notes	E_{sp} [kJ/kg]	Q [kJ/s]	t [°C]	Dissolved Solids (%)	water (m3/hr)
Chem Prep						
w15b	FW after passing through white liquor heater to WW	83	20	20.0		0.8
w16b	FW consumed in chem. prep	8		2.0		128.7
w17b	FW heated and sent to WW tank in chem prep	209	2811	50.0		49.1
w18b	HW consumed in chem. prep	272	425	65.0		5.8
w23b- Total	water in chemicals entering C.P		226			3.1
w27b	FW going to chiller in CP	0	0	0.0	0.0	0.0
Steam Plant						
w19b	FW consumed in boiler condensate tank	8	451	2.0		195.6
w21b	FW consumed in condensate stripper	8	184	2.0		79.7
w25b	water entering with HOG	9	56	5.5	0.0	3.1
Recaust						
w20b	FW consumed in recausticizing	8	86	2.0		37.3
w2b	FW Heated through green liquor cooler	398	2875	95.0		26.0
w4b	FW Heated through dust vent scrubber exchanger	377		90.0		5.0
w26b	water in NaOH in recaust	51	0	20.0	50.0	0.0
Effluents						
Digester						
E6b	NCG'S To turp recovery	411	518	98.0		4.7
Washing						
E1b	Effluent in washing leaving to sewers	313	2041	75.8	1.6	23.5
Bleaching						
E7b	Acidic effluents in bleaching	267	21821	64.0	0.4	298.3
E8b	Alkali effluents in bleaching	313	36697	75.1	0.5	430.6
Pulp Machine						
E2b	Effluent in pulp machine leaving to sewers	302	5209	72.4	0.1	67.8
E9b	Water leaving with pulp	132	1398	90.0	1.6	2.1
E10b	overflows in P.M	322	3163	77.4	0.1	36.4

E14b	water in flue gases from dryer	763	34388	74.8	0.0	43.3
Stream	Notes	E_{sp} [kJ/kg]	Q [kJ/s]	t [°C]	Dissolved Solids (%)	water (m3/hr)
E14b	water from pump seal, flushing, cooling	8	329	1.9	0.0	150.0
Evaporators						
E3b	Effluent in evaporators leaving to sewers	237	4085	56.7	0.0	63.0
C10b	Bl cd going to sewer	272	2492	65.2		33.6
	BL CD going to Recaust A hot water tank	272		65.2		169.9
Chem Prep						
E4b	Effluent in chem. prep leaving to sewers	57	38	13.9	1.8	2.4
E11Ba	from R8	74	2470	17.8	0.00	119.79
E11Bc	saltcake	203	441	57.7	20.82	6.32
E18b	water leaving C.P	210	289	50.3	0.0	5.2
E18ba	water leaving C.P through floor trap	204	460	49.0	0.0	8.5
Steam Plant						
E16BRB	FROM RB	820	89662	253.	0.00	86.72
E16BBB	FROM BB	918	14837	149.	0.00	15.09
BD1	Blowdown from RB 5	1105	3419	253.	0.00	14.00
BD2	Blowdown from PB 4	1109		254.		2.61
	Stripper condensate to bleaching A	277		66.2		88.21
Recasut						
E5b	Effluent in Recaustification leaving to sewers	307	4198	74.3	0.1	61.5
E15b	water in stack losses from lime kiln	1056	2319	250.	0.0	9.5
E15ba	vent from classifier in recaust	411	92	98.0	0.0	0.8
E17b	overflow in white liquor storage tank	0	0	0.0	0.0	0.0
	lime cooling water	49	292	11.9		21.4
	seal water	8	37	2.0		15.9
	water from lime dust slurry tank	288	7	81.8	0.02	0.07

List of heat exchangers – Line B		
Heat Ex.	location	name
HX1	Digester	WBL Cooler
HX2	Digester	flashed steam condenser
HX3	Pulp Machine	Shower water heaeter
HX4	Evaporators	Surface condenser
HX5	Water Production	direct Condenser
HX6	Chemical preparation	Indirect contact cooler
HX7	Chemical preparation	Surface Condenser
HX8	Chemical preparation	Chiller
HX9	Recausticizing	Green Liquor Cooler
HX10	Recausticizing	Dust Vent Scrubber Exchanger
HX11	Water Production	Bleach Heater
HX12	Chemical preparation	White liquor Cooler

ANNEXE 2 – Guidelines for constraint analysis in a Kraft mill

2-1 The complete list of all direct injection steam constraints

2-2 water system data for retrofit and grassroot

2-3 Lists of streams used to build the initial composite curves +

2-4 Actual heating requirement

2-5 Information and equations used to evaluate the total area and capital cost calculations.

2-6 water no water composite curves

2-7 integrated. separated composite curves

2-8 List of streams for the NIM projects

2-9 List of effluents after refinement

2-1 The complete list of all direct injection steam constraints Line A and Line B

Steam constraints – Line B			
Steam Type	Consumer	Heat load (MW)	ID
Low Pressure	Steaming Vessel	8.47	1
	Evaporator effect 2	18.79	56
	Evaporator effect 1	21.28	57
	Machine Showers	16.67	33
	Machine Lazy Showers	5.13	34
	Deaerator*	10.10	62
	Desuperheating	2.94	67
	Bleaching injection	5.35	21,22,23,24
Medium Pressure	O2 Delignification	3.82	5
	Lower Liquor Heater	2.30	2
	Upper Liquor Heater	3.54	3
	Dryer	28.22	35
	Desuperheating	3.46	66
High Pressure	Turbine MP	4.81	68
	Turbine LP	18.95	69
	Soot blowing	13.11	Sb-a
Total		**166.94**	

Steam constraints – Line B			
Steam Type	Consumer	Heat load (MW)	ID
Low Pressure	Steaming Vessel	16.86	1
	Concentrator	30.02	47
	Evaporator effect 1	19.98	48
	Machine Showers	10.96	29
	White water tank	5.81	30
	Deaerator	20.30	53
	Desuperheating	0.94	57
	Reboiler in R8	3.71	68
		5.66	21,22,23,24
Medium Pressure	Lower Liquor Heater	6.11	2
	Upper Liquor Heater	10.29	3
	Dryer	29.25	31
	Desuperheating	2.39	58
High Pressure	Turbine MP	3.21	63
	Turbine LP	23.80	64
	Soot blowing	16.83	Sb-b
Total		**206.11**	

2-2 water system data for retrofit and grassroot line A and line B

Grarssroot cold streams for water system:

	Water type	Department	type	#	T_{in} (°C)	T_{out} (°C)	Q (MW)	Flow t/d)
line B	Hot water	Washing	C	1	2	65.0	6.95	2282
		Bleaching	C	2	2	65.0	27.96	9177
		Chemical Preperation 1	C	3	2	65.0	0.41	135
		Recausticizing 1	C	4	2	95.0	6.07	1343
		Recausticizing 2	C	5	2	90.0	0.52	123
		Hot Water to Tank A	C	17	2	60.0	5.61	2000
	Warm Water	Washing	C	11	2	50.0	5.62	2426
		Machine	C	12	2	65.0	6.55	2150
		Bleaching	C	13	2	50.0	3.30	1423
		Chemical Preperation 2	C	14	2	20.0	0.02	20
		Warm Water to Tank A	C	19	2	49.7	12.26	5327
Line A	Hot water	Bleaching	C	6	2	80.0	25.19	6663
		Machine	C	7	2	80.0	14.47	3827
		Recausticizing 1	C	8	2	80.0	8.53	2256
		Recausticizing 2	C	9	2	90.0	2.59	608
		From Hot Water Tank B to blech	C	18	60	80.0	1.95	2000
	Warm Water	Washing	C	15	2	56.8	3.75	1418
		Bleaching	C	16	2	56.8	3.76	1419
		From Warm Water Tank B to bleach	C	20	50	80.0	7.88	5327

It is important to note that the hot streams used for the water system are the same as the hot streams extracted for the retrofit system except for the non isothermal mixing streams in water tanks. The main difference revolves around the extraction of cold water streams. The hot streams are presented with the retrofit stream extraction below.

Retrofit hot streams for water system

Notes	Extraction point	type	#	T_{in} (°C)	T_{out} (°C)	Q (MW)	Flow rate (t/d)
Process Hot Streams Line B	Coldblow cooler	H	1	100.8	70.0	13.46	9419
	Surface Condenser	H	3	56.7	56.6	40.87	1492
	C.P Surface Condenser	H	5	81.3	19.6	0.01	16
	C.p Indirect Contact Condenser	H	7	100.5	52.6	2.69	178
	Flashed Steam Condenser	H	12	100.0	98.0	2.86	109
	Green liquor cooler	H	20	114.7	94.0	6.07	6546
	Shower water heater	H	22	146.6	65.0	1.57	2150
	Dust vent scrubber exchanger	H	24	100.4	98.1	0.52	20
	White liquor Cooler	H	26	98.0	30.0	0.02	6
Process Hot Streams Line A	Surface Condensers 1,	H	28	59.4	58.4	32.72	1265
	Surface Condensers 2	H	30	78.8	75.8	4.41	164
	Cold Blow Cooler	H	36	95.7	90.0	2.79	10307
	Flashed Steam Condenser	H	38	99.0	94.9	6.46	245
	Green Liquor Cooler	H	40	154.7	95.0	8.53	3510
	Dust Vent Scrubber exchanger	H	49	100.4	98.0	2.59	99
Water Hot Streams Line B	Warm Water Tank B	H	9	50.0	49.7	0.02	1163
	Warm Water Tank B	H	10	60.0	49.7	2.41	4802
	NIMP	H	15	60.0	50.0	0.28	400
	Hot Water Tank B	H	17	84.3	60.0	2.86	2413
Water Hot Streams Line A	Warm Water Tank A	H	33	67.0	56.8	5.87	11807
	Hot Water Tank A	H	42	80.0	68.8	1.35	2467
	Hot Water Tank A (17)	H	44	70.0	68.8	0.61	10039
	Hot Water Tank A (17)	H	43	80.0	68.8	0.82	1500

Retrofit Cold streams for water system

Notes	Extraction point	type	T_{in} (°C)	T_{out} (°C)	Q (MW)	Flow rate (t/d)
Water Cold Streams Line B	Coldblow cooler	C	2	60	13.46	4802
	Surface Condenser	C	2	47	40.87	18817
	C.P Surface Condenser	C	2	50	0.01	2
	C.pICC	C	2	50	2.69	1161
	Warm Water Tank B	C	47	49.7	2.43	18817
	Flashed steam cond.	C	50	55	2.86	11757
	Direct Condenser	C	49.7	84.3	4.4	2413
	NIMP	C	49.7	50	0.28	17180
	Hot Water Tank B	C	55	60	2.86	11757
	Bleach HX B	C	60	65	2.82	11594
	Green liquor cooler	C	2	95	6.07	1343
	Shower water heater	C	50	65	1.57	2150
	DVSE	C	2	90	0.52	123
	White liquor Cooler	C	2	20	0.02	20
Water Cold Streams Line A	Surface Condensers 1	C	2	59.3	32.72	11807
	Surface Condensers 2	C	59.3	67	4.41	11807
	Direct Condenser	C	2	2	0	1520
	Warm Water Tank A	C	2	56.8	4.02	1520
	Warm Water Tank A	C	49.7	56.8	1.85	5327
	Cold Blow Cooler	C	56.8	80	2.79	2467
	Flashed Steam Cond	C	56.8	70	6.46	10039
	Green Liquor Cooler	C	2	80	8.53	2256
	Hot Water Tank A	C	56.8	68.8	1.93	3312
	Hot Water Tank A	C	60	68.8	0.85	2000
	Brown HX A	C	68.8	80	3.24	5900
	Bleach HX A	C	68.8	80	7.37	13418
	DVSE	C	2	90	2.59	608

2-3 Lists of streams used to build the initial composite curves + actual heating requirement

Line A

Departmen	Definition	Typ	Tin	Tout	Q	Flow rates
Digester	LP into Steaming Vessel	C	169.0	170.0	8.47	331
	Liquor heated using MP	C	145.6	154.0	2.30	6085
	Liquor heated using MP	C	138.3	151.8	3.54	5819
	cooling effluent	H	94.9	30.0	0.77	245
washing	MP steam injection	C	201.0	202.0	3.82	135
	Cooling effluent	H	76.4	30.0	0.67	306
	Cooling effluents before	H	77.3	30.0	0.28	123
bleaching	LP injected in pulp line	C	169.0	170.0	3.61	127
	LP injected in pulp line	C	169.0	170.0	3.92	139
	LP injected in pulp line	C	169.0	170.0	3.83	134
	LP injected in pulp line	C	169.0	170.0	2.92	101
	Effluent Cooling	H	72.4	30.0	34.10	16648
	Effluent Cooling	H	69.5	30.0	18.60	9763
	Cooling overflow	H	82.6	30.0	1.74	686
	Cooling overflow	H	71.0	30.0	1.61	814
Machine	LP injected in lazy showers	C	169.0	170.0	16.67	575
	LP injected in showers	C	169.0	170.0	5.13	180
	MP used in the dryer	C	201.0	202.0	28.22	1199
	air leaving air-air	H	93.4	73.8	1.08	3811
	Cold glycol	C	61.2	70.0	1.08	4073
	Hot glycol	H	70.0	61.2	1.08	4073
	Air entering to machine	C	-8.0	25.0	1.08	2863
	Hot exhaust air leaving	H	132.0	93.4	2.17	3811
	Air from air-glycol heater	C	25.0	90.0	2.17	2863
	Exhaust	H	73.8	30.0	2.42	3811
Evaporator	BL heating in effect 2	C	113.5	125.6	18.79	4657
	BL heating in effect 1	C	125.6	127.5	21.28	4046
	Steam from 4th effect	H	91.5	88.5	17.24	2624
	liquor from 6th effect to 5th	C	60.4	60.5	17.24	650
Steam	Make up water	C	169.0	170.0	16.07	5516
	Air heated using LP steam	C	45.0	80.0	3.16	7750
	Water for MP	C	137.5	202.0	3.46	133
	Water for LP	C	137.5	170.0	2.94	115
	MP turbine	C	271.6	400.0	4.81	1863
	LP turbine	C	195.5	400.0	18.95	4582
	blowdown - RB	H	253.7	30.0	2.36	208
	sootblowing	C	399.0	399.1	13.11	404
Recaust	Energy in vent gases	H	98.0	30.0	0.33	99
	Cooling stack gases - -kiln	H	250.0	100.0	1.38	184
	Effluent to sewer	H	226.3	30.0	2.70	285

Line B

Department	# Stream	Type	Tin	Tout	Q	Flow rate(
Digester	LP into Steaming Vessel	C	169.0	170.0	16.86	654
	Liquor heated using MP	C	170.3	187.0	6.11	7983
	Liquor heated using MP	C	154.3	187.0	10.29	6908
	cooling effluent	H	98.0	30.0	0.36	109
Washing	Effluents out of washing	H	75.8	30.0	1.24	563
Bleaching	LP injected in pulp line	C	169.0	170.0	7.36	263
	LP injected in pulp line	C	169.0	170.0	4.20	149
	LP injected in pulp line	C	169.0	170.0	3.37	119
	LP injected in pulp line	C	169.0	170.0	2.75	96
	Energy from acid effluent	H	64.0	30.0	11.64	7074
	Energy from alkaline effluent	H	75.1	30.0	22.11	10132
Machine	LP injected in showers	C	169	170.0	10.96	400
	LP injected in white water	C	169.0	170.0	5.55	200
	MP used in the dryer	C	201.0	202.0	29.25	1147
	air leaving air-air heat	H	93.7	74.8	1.08	3891
	Cold glycol	C	70.9	75.0	1.08	8640
	Hot glycol	H	75.0	70.9	1.08	8640
	Air entering to machine room	C	-8.0	25.0	1.08	2863
	Hot exhaust air leaving dryer	H	131.1	93.7	2.17	3891
	Air from air-glycol heater	C	25.0	90.0	2.17	2863
	Exhaust Air	H	74.8	30.0	2.56	3891
	Effluents out of washing	H	72.4	30.0	3.06	1491
Evaporator	BL heating in concentrator	C	97.7	130.1	30.02	3451
	BL heating in effect 1	C	88.9	105.3	19.98	4942
	Last effect condensate to	H	65.2	30.0	2.83	1665
	Condensate from S.C	H	56.7	30.0	1.92	1492
Steam	Make up water	C	169.0	170.0	36.40	10135
	Air heated using LP steam	C	35.0	80.0	7.59	7459
	Water for MP desuperheating	C	138.0	186.1	0.94	109
	Water for LP desuperheating	C	137.5	147.9	2.39	136
	blowdown IN RB	H	253.7	30.0	3.03	267
	blowdown in PB	H	137.5	30.0	0.33	63
	Energy from PB stack gases	H	149.7	100.0	0.97	1392
	MP turbine	C	268.1	400.0	3.21	1200
	LP turbine	C	188.5	400.0	23.80	5522
	sootblowing	C	399.0	399.1	16.83	519
Recaust	Energy in vent gases	H	72.8	30.0	0.02	16
	Cooling stack gases from kiln	H	250.0	100.0	0.91	188
	Effluent to sewer	H	77.3	30.0	1.67	1166
Chem. Prep	R8 chemicals in the reboiler	C	141.1	147.9	3.71	30102
	Water cooling from condenser	H	66.1	30.0	0.17	122

2-4 Actual heating requirement (AHR)

The actual heating requirement can be calculated either by considering the energy in the steam produced minus the condensate that returns in the condensate collection tanks. The other way is to calculate the steam consumed at every location. Some of that steam is directly injected and consumed completely while in other case it is used in indirect heat exchangers. The total actual consumption was evaluated using the second method for line A and line B. This is done by adding the steam that is being used as constraints and non constrains.

Line A:

Constrained steam =166.94

Non constrained steam = 28.71

Actual heating requirement = 196 MW

Line B:

Constrained steam = 206.11 MW

Non constrained steam = 42.87 MW

Actual heating requirement = 249 MW

<u>2-5 Information and equations used to evaluate the total area and capital cost calculations.</u>

The information was obtained from Aspen energy analyzer help menu.

Total Area:

$$A_{\text{network}} = \sum_{k=1}^{n \text{ intervals}} \frac{1}{\Delta T_{IM,k}} \left(\sum_{i}^{\text{hot streams}} \frac{q_i}{h_i} + \sum_{i}^{\text{cold streams}} \frac{q_i}{h_i} \right)$$

ΔT_{lm} is the logarithmic temperature difference between the hot and cold streams
q_i is the enthalpy change through a specific temperature interval
h_i is the heat transfer coefficient

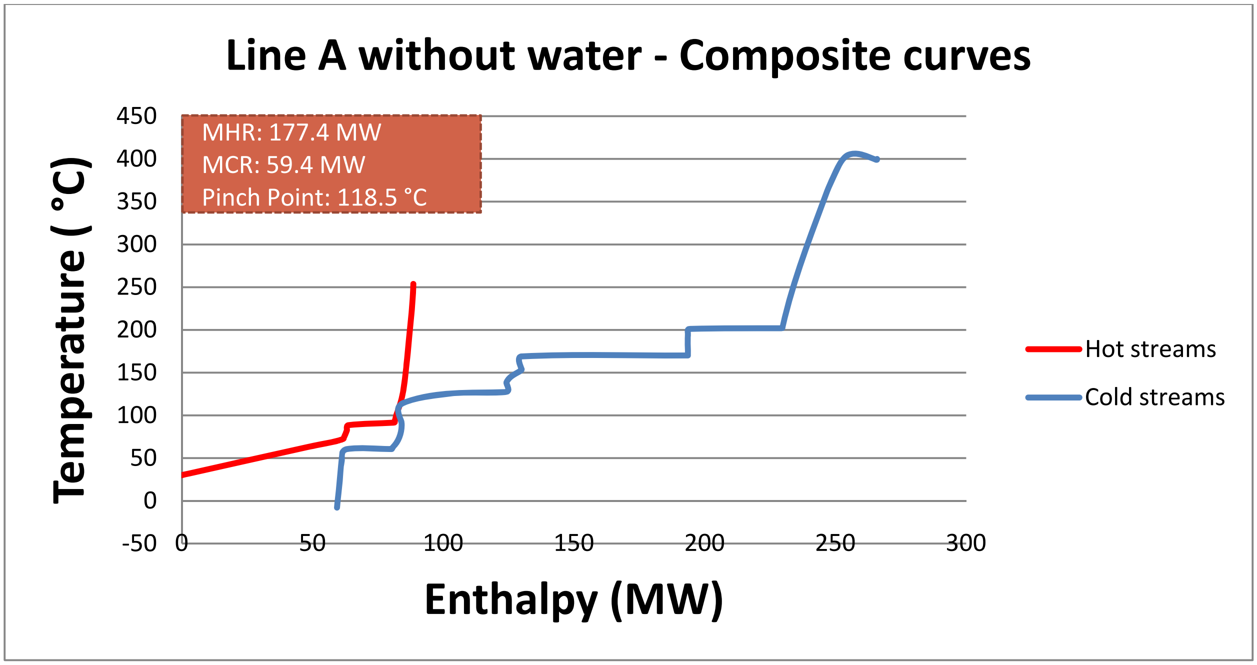
Capital Cost Index:

$$\text{Capital Costs Index} = a + b\left(\frac{A_{exchange}}{Shell}\right)^{c} \cdot Shell$$

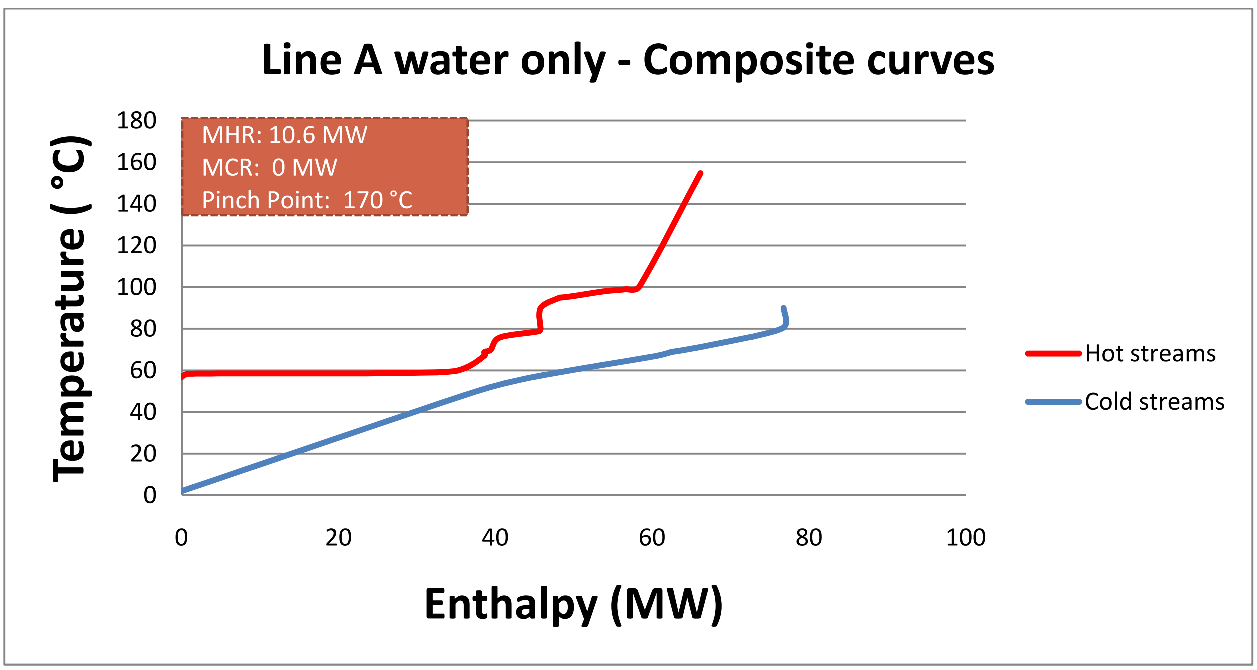
a = 10000 , b = 800 , c = 0.8
$A_{exchange}$ = Total area (m^2)
Shell is the number of shells in a shell and tube heat exchanger

2-6 integrated water or separated water system composite curves

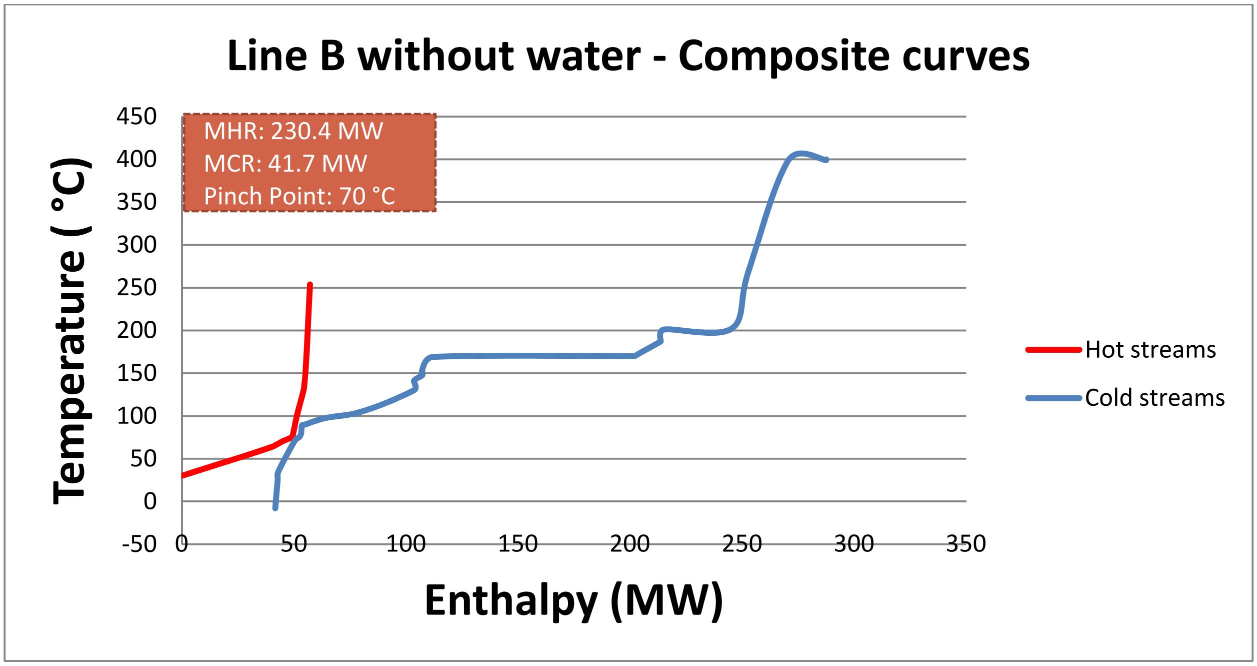
Line A – without water (Separated)



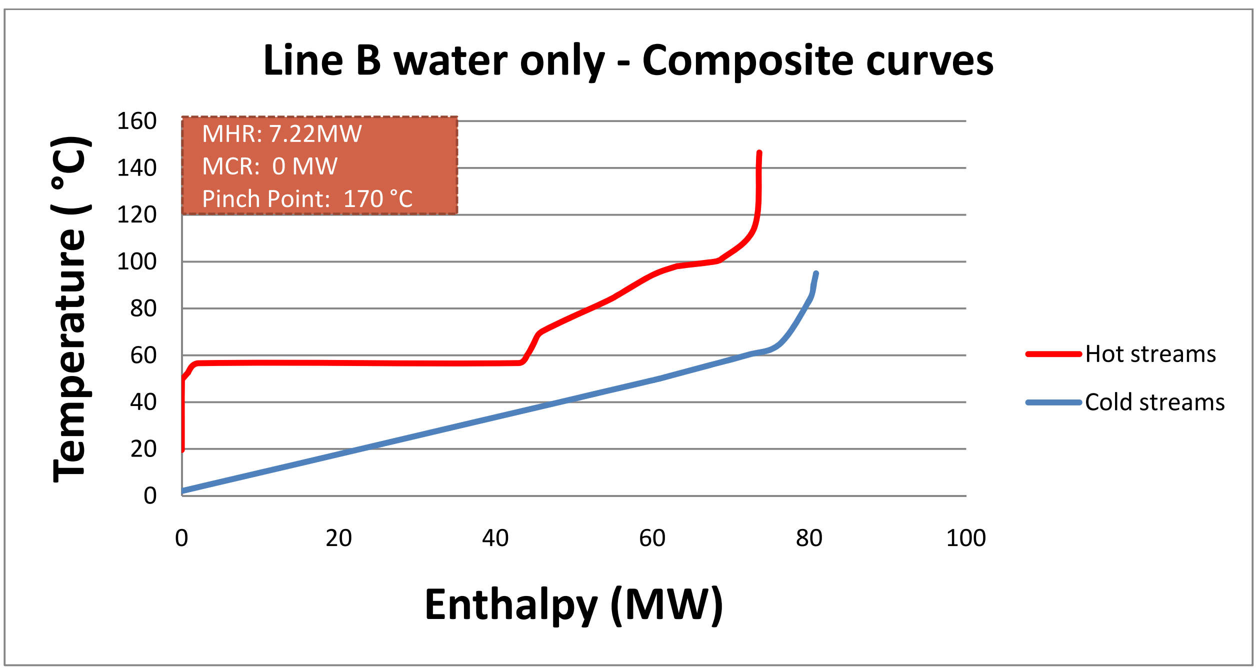
Line A – water only



Line B – without water (Separated)

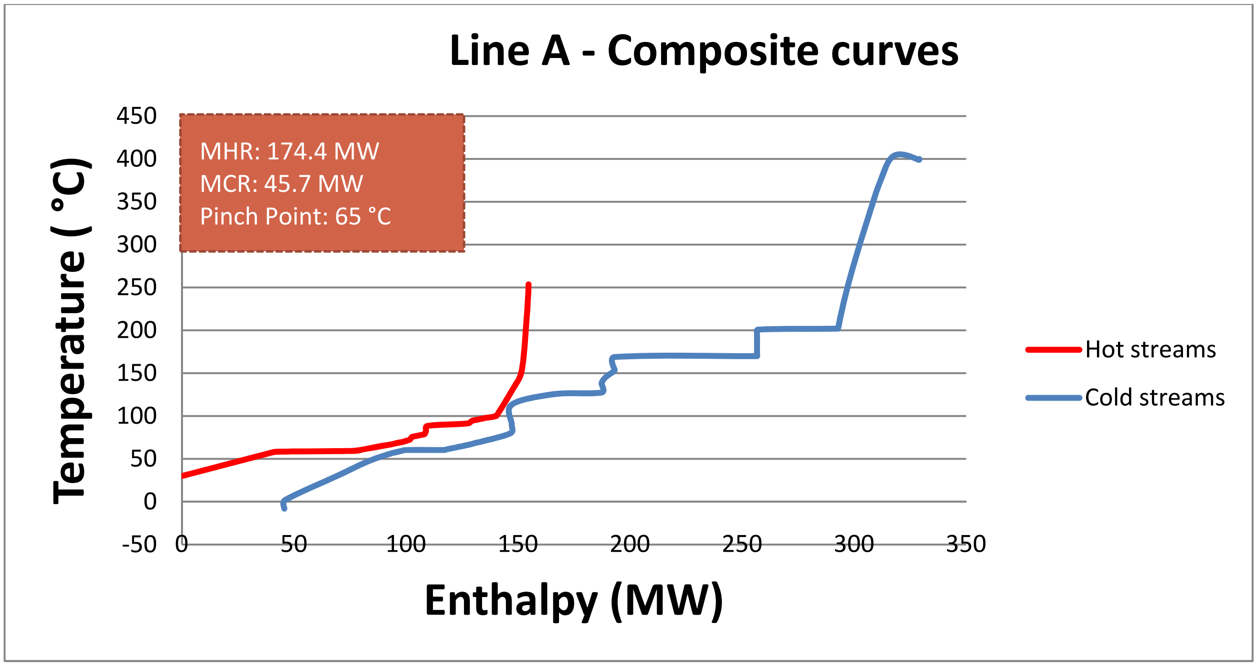


Line B – water only

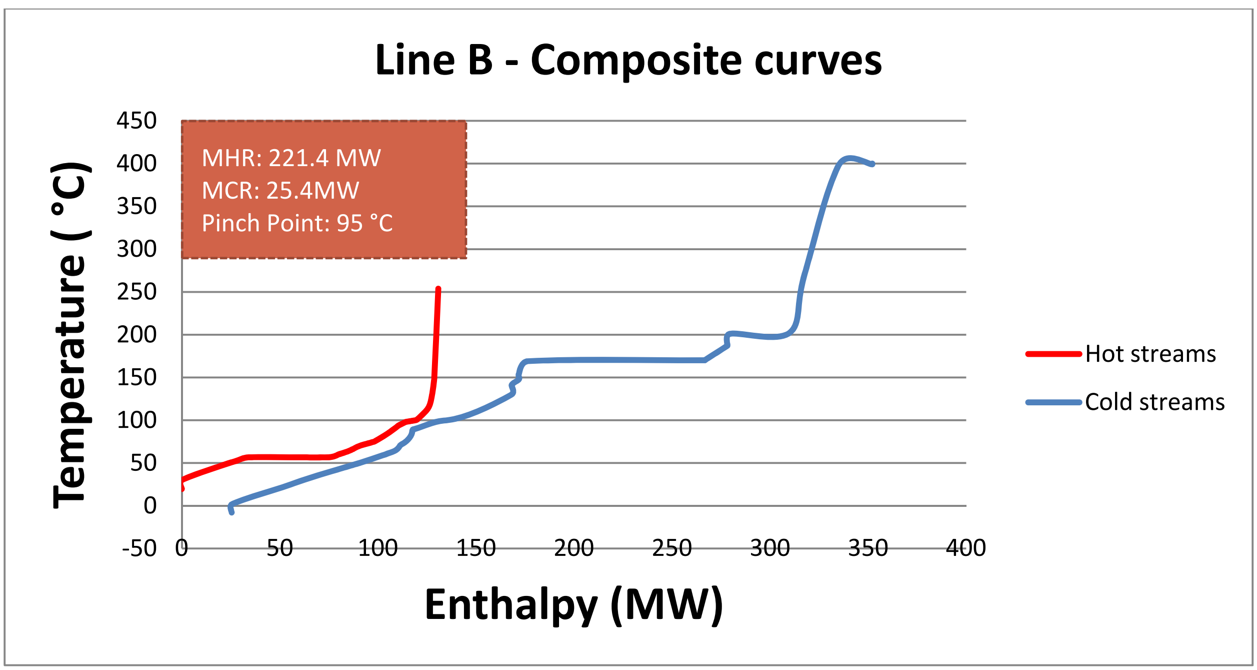


2-7 integrated vs. separated lines composite curves

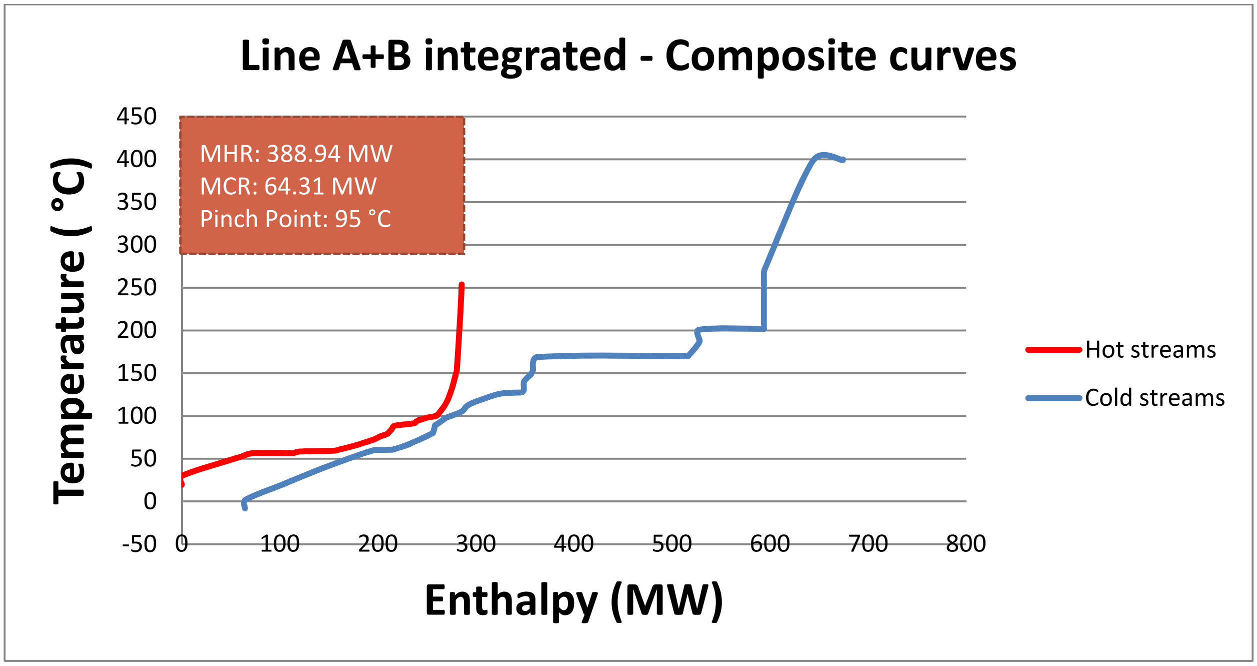
Line A



Line B



Line A + B integrated



2-8 List of streams for the NIM projects

Line A

Cold Streams						
Department	Definition	Tin (°C)	Tout (°C)	mCp (KJ/ C h)	load (MW)	Flow (t/d)
Bleaching	Injection 2-1	2	60	1.37E+05	2.21	791
Bleaching	Injection 4-1	2	60	1.37E+05	2.21	791
Steam Plant	Heating make up	31.0	95.0	3.38E+05	6.00	1915
Bleaching	Injection 1-3	57.4	82	1.74E+05	1.19	1000
Bleaching	Injection 2-2	66.7	88	7.54E+05	4.46	4325
Bleaching	Injection 2-3	69.6	86.5	3.68E+05	1.73	2110
Bleaching	Injection 3-2	69.6	86.5	6.36E+05	2.98	3645
Bleaching	Injection 3-3	63.6	88	5.44E+05	3.69	3122
Bleaching	Injection 4-1	63.6	88	3.91E+05	2.65	2246
Bleaching	Injection 4-2	57.4	82	1.01E+05	0.69	5371

Line A

Cold Streams						
Department	Definition	Tin (°C)	Tout (°C)	mCp (KJ/ C h)	Load (MW)	Flow (t/d)
Steam Plant	Makeup	31.0	95.0	9.06E+05	16.10	4692
Machine	FW to WW tank	2.0	77.4	1.53E+05	3.20	718
Bleaching	Injection 1-2	70.0	83.1	7.24E+05	2.63	4149
Bleaching	Injection 1-2	65.0	84.0	5.68E+05	3.00	3264
Bleaching	Injection 2-1	70.0	83.1	8.84E+05	3.22	5035
Bleaching	Injection 2-2	71.6	84.0	3.92E+05	1.35	2248
Bleaching	Injection 3-1	71.6	84.0	4.53E+05	1.56	2612
Bleaching	Injection 3-2	69.4	79.0	6.93E+05	1.85	3981
Bleaching	Injection 4-2	69.4	79.0	1.01E+06	2.69	5798
Bleaching	Injection 4-3	65.0	79.0	5.68E+05	2.21	2117
Washing	WW-dilution conveyer	50.0	74.0	3.72E+05	2.48	2141

2-9 list of effluents after refinement

Effluents:

The following table will contain a description of the effluents after refinement. Screening of the significant effluents that have the potential to be used as heat sources will be carried based on the energy load in the streams. The criteria for evaluation will include the heat load, consistency and temperature levels.

Line A Department	Effluent location	Dissolved Solids (%)	Consistency (%)	T_{in} (°C)	T_{out} (°C)	Heat load (MW)
Digester	Flash steam condenser	0	0	95.0	30.0	0.77
Washing	Dum Drainer effluent	1.9	1.2	76.4	30.0	0.67
Bleaching	Acidic effluent	0.3	0	72.4	30.0	34.10
	Alkaline effluent	0.5	0	69.5	30.0	18.60
	Wash press filtrate tank	0.9	0	82.6	30.0	1.74
	contaminated cond. tank	0	0	70.9	30.0	1.61
Machine	Exhaust Air	0	0	73.8	30.0	2.42
	Reject tank	0	1	56.7	30.0	0.65
Steam Plant	Stack gases from RB1	0	0	252.8	134.3	17.43
	Blowdownfrom RB1	0	0	253.7	30.0	2.36
Recasut.	Vent gases	0	0	98.0	30.0	0.32
	Lime kiln stack	0.04	0.7	250.0	100.0	1.38
	Effluents to sewer	0.02	4.1	78.0	30.0	0.64

Effluents – Line B

Line B						
Department	Effluent location	Dissolved Solids (%)	Consistency (%)	T_{in} (°C)	T_{out} (°C)	Heat load (MW)
Digester	Flash steam condenser	0	0	98.0	30.0	0.36
Washing	Effluents to sewer	1.6	0.1	75.8	30.0	1.24
Bleaching	Acidic effluent	0.4	0	64.0	30.0	11.64
	Alkaline effluent	0.5	0	75.1	30.0	22.11
Machine	Exhaust Air	0	0	74.8	30.0	2.5
	Reject tank	0.1	0.4	72.4	30.0	3.06
Evaporators	Surface condenser	0	0	65.2	30.0	2.83
	Condensate to sewer	0	0	56.7	30.0	1.92
Steam Plant	Stack gases from RB5	0	0	253.7	133.0	15.94
	Stack gases from PB2	0	0	149.7	100.0	0.97
	Blowdownfrom RB5	0	0	253.7	30.0	3.03
	Blowdownfrom PB2	0	0	137.5	30.0	0.33
Recasut	Lime kiln stack	0.04	3	250.0	100.0	0.91
	Effluents to sewer	0.2	1.3	77.3	30.0	1.67

ANNEXE 3 – Heat exchanger network and energy saving projects

3-1 List of streams used for building the heat exchanger network

3-2 Economic analysis data

3-3 List of modified heat exchanger networks and projects for line A

3-4 List of modified heat exchanger networks and projects for line B

3-1 <u>List of streams used for building the heat exchanger network</u>

Line A retrofit

Hot stream							
Department	Definition	Tin (°C)	Tout (°C)	mCp (KJ/C h)	load (MW)	Flow (t/d)	Is the stream being used?
BELOW							
Evap.	Vapour to SC 1	59.4	58.4	1.18E+08	32.72	1265	SC 1
Water prod	WW-SC	67.0	56.7	2.07E+06	5.87	11807	NO-WW tank A
Bleaching	Alkaline effluent	69.5	30.0	1.69E+06	18.60	9763	NO
Water Prod	HW-FSC	70.0	68.7	1.76E+06	0.61	10039	NO-HW tank A
Machine	Hot Glycol	70.0	61.2	4.45E+05	1.08	4073	Air Heater
BELOW AND ABOVE							
Bleaching	Acidic effluent	72.4	30.0	2.89E+06	34.10	16648	NO
Machine	Exhaust Air	73.7	30.0	1.99E+05	2.42	3811	NO
Washing	Drainer effluent	76.4	30.0	5.22E+04	0.67	306	NO
Recasut	Effluents - sewer	78.0	30.0	4.80E+04	0.64	285	NO
WaterProd	HW- Brown HX	79.9	68.7	2.64E+05	0.82	1500	NO-HW tank A
WaterProd	HW-blow cooler	80.0	68.7	4.34E+05	1.35	2467	NO-HW tank A
Recaust	Classifier gases	97.9	30.0	1.73E+04	0.33	99	NO
SteamPlan	Blowdown- RB	253	30	3.80E+04	2.36	208	NO
ABOVE							
Evap.	Vapour - S.C 2	78.8	75.8	5.29E+06	4.41	164	S.C2
Evap.	Steam-4theffect	91.5	88.5	2.07E+07	17.24	2624	Black Liquor Heater
Machine	Hot Air	93.3	73.7	2.00E+05	1.08	3811	Glycol heater
Digester	WBL	95.7	90.0	1.76E+06	2.79	10307	CBC
Digester	Vapour	99.0	94.9	5.67E+06	6.46	245	FSC
Recasut	gases Classifier	100	98.0	3.88E+06	2.59	99	DVSE
Machine	Hot air	131	93.3	2.02E+05	2.17	3811	Air-Air Heater
Recasut	Green liquor	154	95.0	5.14E+05	8.53	3510	GLC
Recasut	LimeKiln Stack	250	100	3.31E+04	1.38	184	NO
SteamPlan	Stackgases- RB	252	134	5.30E+05	17.43	10257	No

Cold Streams

Department	Definition	Tin (°C)	Tout (°C)	mCp (KJ/ C h)	Heat load (MW)	Flow (t/d)	Is the stream being used?
BELOW							
Evaporators	Fresh Water	2.0	59.3	2.06E+06	32.72	11807	S.C 1
Water Prod.	WW Tank B to A	49.7	56.8	9.32E+05	1.85	5327	NO-WW tank A
Water Prod.	WW-Condenser A to WW Tank A	2.0	56.8	2.64E+05	4.02	1520	NO-WW tank A
Machine	Fresh Air	-8.0	25.0	1.18E+05	1.08	2863	Air Heater-Glycol
Bleaching	Injection 2-1	2	60	1.37E+05	2.21	791	No- new project
Bleaching	Injection 4-1	2	60	1.37E+05	2.21	791	No- new project
BELOW AND ABOVE							
Steam Plant	Heating make up	31.0	95.0	3.38E+05	6.00	1915	No- new project
Recaust.	Fresh Water	2.0	90.0	1.06E+05	2.59	608	DVSE
Machine	Dryer Air	25.0	90.0	1.20E+05	2.17	2863	Air-Air Heater
Digester	Warm water	56.8	80.0	4.33E+05	2.79	2467	Cold Blow Cooler
Recaust.	Fresh Water	2.0	80.0	3.94E+05	8.53	2256	GLC
Steam Plant	Boiler Air	45.0	80.0	3.25E+05	3.16	7750	Air Heater
Digester	WW to HW	56.8	70.0	1.76E+06	6.46	10039	Flashed Steam Condenser
Water Prod.	WW - WW Tank A to HW tank A	56.8	68.8	5.81E+05	1.93	3312	NO-Hot Water tank A
Evaporators	Fresh Water	59.3	67.0	2.06E+06	4.41	11807	S.C 2
ABOVE							
Water Prod.	Water- HW Tank	68.8	80.0	1.04E+06	3.24	5900	Brown HX A
Water Prod.	Water – HW Tank	68.8	80.0	2.37E+06	7.37	13418	Bleach HX A
Water Prod.	HW Tank B to tank A	60.0	68.8	3.51E+05	0.85	2000	NO-HW tank A
Machine	Glycol	61.2	70.0	4.45E+05	1.08	4073	Glycol heater
Evaporators	B liquor-6th effect	60.4	60.5	6.21E+08	17.24	650	Black Liquor Heater
Bleaching	Injection 1-3	57.4	82	1.74E+05	1.19	1000	No- new project
Bleaching	Injection 2-2	66.7	88	7.54E+05	4.46	4325	No- new project
Bleaching	Injection 2-3	69.6	86.5	3.68E+05	1.73	2110	No- new project
Bleaching	Injection 3-2	69.6	86.5	6.36E+05	2.98	3645	No- new project
Bleaching	Injection 3-3	63.6	88	5.44E+05	3.69	3122	No- new project
Bleaching	Injection 4-1	63.6	88	3.91E+05	2.65	2246	No- new project
Bleaching	Injection 4-2	57.4	82	1.01E+05	0.69	5371	No- new project

Line A Grassroot

The hot streams are the same except for non isothermal mixing in water tanks. In the grassroot extraction, these streams do not exist. Process .The table below has the cold stream used to build the network:

Cold Streams - Water						
Department	Destination	Tin (°C)	Tout (°C)	mCp (KJ/ C h)	Heat load (MW)	Flow (t/d)
BELOW						
N.A	Washing	2	56.8	2.46E+05	3.75	1418
N.A	Bleaching	2	56.8	2.47E+05	3.76	1419
Bleaching	Injection 2-1	2	60	1.37E+05	2.21	791
Bleaching	Injection 4-1	2	60	1.37E+05	2.21	791
BELOW AND ABOVE						
N.A	Bleaching	2	80	1.16E+06	25.19	6663
N.A	Machine	2	80	6.68E+05	14.47	3827
N.A	Recausticizing 1	2	80	3.94E+05	8.53	2256
N.A	Recausticizing 2	2	90	1.06E+05	2.59	608
N.A	Warm water tank B to bleaching	50	80	9.46E+05	7.88	5327
Steam Plant	Make up water	31	95	3.38E+05	6	1731
ABOVE						
N.A	HW tank B to bleaching	60	80	3.51E+05	1.95	2000
Bleaching	Injection 1-3	57.4	82.0	1.74E+05	1.2	1000
Bleaching	Injection 2-2	66.7	88.0	7.54E+05	4.5	4325
Bleaching	Injection 2-3	69.6	86.5	3.68E+05	1.7	2110
Bleaching	Injection 3-2	69.6	86.5	6.36E+05	3.0	3645
Bleaching	Injection 3-3	63.6	88.0	5.44E+05	3.7	3122
Bleaching	Injection 4-1	63.6	88.0	3.91E+05	2.7	2246
Bleaching	Injection 4-2	57.4	82.0	1.01E+05	0.7	5371

Line B Retrofit

Hot stream							
Department	Definition	Tin (°C)	Tout (°C)	mCp (KJ/ C h)	Load (MW)	Flow (t/d)	Is the stream being used?
BELOW							
Evaporator	Condensate - S.C	56.7	30	2.59E+05	1.92	1492	NO
Evaporator	vapour form S.C	56.7	56.6	1.47E+09	40.87	1492	YES – S.C
Water Prod	WW from S.C	50	49.6	2.03E+05	0.02	1163	NO-NIM WW tank B
BELOW AND ABOVE							
Washing	Effluent out	75.8	30	9.75E+04	1.24	563	NO
Bleaching	Acid effluent	64	30	1.23E+06	11.64	7074	NO
Bleaching	Alkaline effluent	75.1	30	1.76E+06	22.11	1013 2	NO
Machine	Exhaust Air	74.8	30	2.06E+05	2.56	3891	NO
Machine	Effluent out	72.4	30	2.60E+05	3.06	1491	NO
Evaporator	Cond- last effect	65.2	30	2.89E+05	2.83	1665	NO
Steam Plant	Blow down -RB	254	30	4.88E+04	3.03	267	NO
Steam Plant	Blowdown- PB	138	30	1.11E+04	0.33	63	NO
Recaust	Effluent to sewer	77.3	30	1.27E+05	1.67	1166	NO
Chem Prep	R8reactor product	101	52.6	2.02E+05	2.69	178	YES- ICC
Chem Prep	White liquor	98	30	1.06E+03	0.02	6	YES- WL cooler
WaterProd	WW -blow cooler	60	49.6	8.40E+05	2.41	4802	NO- WW tank B
WaterProd	HW - HW tank B	60	49.9	1.01E+05	0.28	400	NO- HW AND WW
ABOVE							
Machine	Air	93.7	74.8	2.06E+05	1.08	3891	YES- Glycol heater
Machine	Glycol	75	70.9	9.48E+05	1.08	8640	YES-Air Heater
Machine	Air-Air Heater	131	93.7	2.09E+05	2.17	3891	YES- Air- Air Heater
Digester	WBL	101	70	1.57E+06	13.4	9419	Cold blow cooler
Chem Prep	CLO2	81.3	19.6	3.05E+02	0.01	16	YES- SC
Digester	Vapour -	100	98	5.15E+06	2.86	109	YES-FSC
Recaust	Green liquor	114	94	1.06E+06	6.07	6546	YES-GLC
Machine	MP cond - dryer	146	65	6.93E+04	1.57	2150	YES-SWH
Recaust	Classifier gases	100	98.1	8.14E+05	0.52	20	YES- DVSE
Water Prod	HW - HW tank B	84.2	60	4.24E+05	2.86	2413	NO-NIM HW tank B
Recaust	Lime Kiln Stack	250	100	2.18E+04	0.91	188	NO
SteamPlant	PB stack gas	149	100	7.03E+04	0.97	1392	NO
SteamPlant	RB stack gas	253	133	4.76E+05	15.9	9442	NO

Cold Streams							
Department	Definition	Tin (°C)	Tout (°C)	mCp (KJ/ C h)	Load (MW)	Flow (t/d)	Is the stream being used?
BELOW							
Machine	Cold Air	-8.0	25.0	1.18E+05	1.08	2863	YES-Air Heater
Recaust	FW	2.0	20.0	3.52E+03	0.02	20	YES- White liquor cooler
BELOW AND ABOVE							
Machine	Warm Air	25.0	90.0	1.20E+05	2.17	2863	YES- Air- Air Heater
Steam Plant	Boiler Air	35.0	80.0	6.07E+05	7.59	7459	Yes-Boiler Air heater
Digester	FW to WW	2.0	60.0	8.36E+05	13.46	4802	YES- Cold blow cooler
Evaporators	FW to WW	2.0	47.0	3.27E+06	40.87	18817	YES-Surface Cond
Chem Prep	FW to WW	2.0	50.0	3.92E+02	0.01	2	YES - ClO2 S.C
Chem Prep	FW to WW	2.0	50.0	2.02E+05	2.69	1161	YES- ICC
Recaust	FW to HW	2.0	95.0	2.35E+05	6.07	1343	YES-Green liquor cooler
Recaust	FW to HW	2.0	90.0	2.14E+04	0.52	123	YES- Dust vent scrubber exchanger
Steam Plant	Makeup	31.0	95.0	9.06E+05	16.10	4692	NEW PROJECT
Machine	FW to ww tank	2.0	77.4	1.53E+05	3.20	718	NEW PROJECT
ABOVE							
Machine	Cold Glycol	70.9	75.0	9.48E+05	1.08	8640	YES- Glycol heater
Digester	WW to HW	50.0	55.0	2.06E+06	2.86	11757	YES- FSC
Machine	WW to SWH	50.0	65.0	3.76E+05	1.57	2150	YES- SWH
Water Prod	WW - S.C	47.0	49.7	3.29E+06	2.43	18817	NO-NIM WW tank B
Water Prod	WW- tank	49.7	50.0	3.00E+06	0.28	17180	NO- HW AND WW
Water Prod	HWfrom FSC	55.0	60.0	2.06E+06	2.86	11757	NO-NIM HW tank B
Water Prod	HWfrom HW tank B	60.0	65.0	2.03E+06	2.82	11594	YES- Bleach Heater
Water Prod	WW to HW tank	49.7	84.3	4.58E+05	4.40	2413	YES-Direct Cond
Bleaching	Injection 1-2	70.0	83.1	7.24E+05	2.63	4149	NEW PROJECT
Bleaching	Injection 1-2	65.0	84.0	5.68E+05	3.00	3264	NEW PROJECT
Bleaching	Injection 2-1	70.0	83.1	8.84E+05	3.22	5035	NEW PROJECT
Bleaching	Injection 2-2	71.6	84.0	3.92E+05	1.35	2248	NEW PROJECT
Bleaching	Injection 3-1	71.6	84.0	4.53E+05	1.56	2612	NEW PROJECT
Bleaching	Injection 3-2	69.4	79.0	6.93E+05	1.85	3981	NEW PROJECT
Bleaching	Injection 4-2	69.4	79.0	1.01E+06	2.69	5798	NEW PROJECT
Bleaching	Injection 4-3	65.0	79.0	5.68E+05	2.21	2117	NEW PROJECT
Washing	WW-dilution conveyer	50.0	74.0	3.72E+05	2.48	2141	NEW PROJECT

3-2 Economic analysis data

In order to evaluate the economical impact of each project, the following data was used:

a. **Cost of steam production from hog and natural gas**

   The information was obtained from the engineer responsible for the steam plant in the mill. The cost of producing 1 tonne of steam from hog and natural gas was evaluated. If steam is produced from hog, it costs 8.11$/tonne of LP steam while for natural gas the price is 17.45 $/ tonne of LP steam.

b. **Cost of electricity**

   This information was also obtained from the mill. It is related to the purchase price of electricity from BC hydro and it equals to .036 $/KWh.

c. **Cost of an installed condensing turbine**

   Based on discussions with the mill engineers, it costs 1.5 M$/MWh for the purchase and installation of a condensing turbine

d. **Net operating cost Savings due to steam reduction**

   The reduction in steam production will have a negative impact on the electricity produced from the turbines and therefore reduce the total operating cost savings due to extra power purchased from the grid. This is evaluated by subtracting the fuel savings ($/yr) from extra power purchased ($/yr). The effect on power reduction was evaluated using Cadsim plus.

e. **Power production from a condensing turbine**

   Software under the title of Turbine steam consumption calculator is used to evaluate power production in a condensing turbine[28]. An efficiency of 80 % is assumed and an exhaust pressure of 0.06 bar. The inlet pressure and temperature are the same as LP steam exiting from the current back pressure turbine. The savings ($/yr) will be obtained from reduction in purchased power.

f. **Indexed price**

   The cost of the heat exchanger was indexed from 1989 to 2011 with index value of 906 and 1522 respectively

g. **Simple Payback:** It is obtained by dividing capital cost by operating savings.

3-3 List of modified heat exchanger networks and projects for line A

Retrofit – low savings

Project	HX area (m2)	Capital Cost (M$)	LP Steam (t/hr)	Steam savings (M$/yr)	Electrical production drop (MW)	electricity purchased (M$/yr)	Net operating savings (M$/YR)
Project 1							
Bleach Heater	521	0.487	12.3	0.863	1.18	0.372	0.491
Project 2							
Brown Heater	239	0.204	5.1	0.358	0.5	0.158	0.200
Below pinch	37	0.028					
Above pinch	202	0.176					
Project 3							
Air Heater	1478	1.107	4.8	0.337	0.46	0.145	0.192
Below pinch	1170	0.834	2.1				
Above pinch	308	0.273	2.7				
Project 4							
Make up Water	537	0.427	9.4	0.663	0.91	0.287	0.376
Below pinch	185	0.112					
Above pinch	352	0.315					
Project 5							
Injection1-2	130	0.113	2.5	0.175	0.24	0.076	0.100
Total	**2905**	**2.338**	**34.1**	**2.396**	**3.29**	**1.038**	**1.359**

Line A - Condensing turbine results – Low savings

Extra electricity (MW)	5.37
Reduction in purchased Electricity (M$/yr)	1.69
Turbine capital cost (M$)	8.06

Retrofit - medium savings

Project	HX area (m2)	Capital Cost (M$)	LP Steam (t/hr)	Steam Savings (M$/yr)	electrical production drop (MW)	electricity purchased (M$/yr)	Net operating savings (M$/YR)
Project 1							
GLC	1292	0.842					
Below pinch	459	0.285					
Above pinch	833	0.557					
DVSE	193	0.123					
Below pinch	86	0.056					
Above pinch	107	0.067					
Bleach Heater	107	0.067	12.3	0.863	1.180	0.372	0.491
Project 2							
FSC	444	0.272					
Above pinch	114	0.071					
Below pinch	330	0.200					
BSH	75	0.057	5.1	0.358	0.500	0.158	0.200
FS	Available						
Blowdown	26	0.022					
Project 3							
Boiler Air HX	1387	1.023	4.8	0.337	0.460	0.145	0.192
Below pinch	1170	0.834	2.1				
Above pinch	217	0.189	2.7				
Project 4							
Makeup Water	418	0.316	9.4	0.663	0.910	0.287	0.376
Below pinch	185	0.112					
Above pinch	233	0.204					
Project 5							
Injection1-2	Available		2.5	0.175	0.241	0.076	0.099
Project 6							
injection 2-1	96	0.061	4.0	0.281	0.386	0.122	0.159
injection 2-2	343	0.307					
injection 2-3	136	0.118					
Project 7							
injection 3-1	103	0.065	4.3	0.302	0.415	0.131	0.171
injection 3-2	210	0.183					
injection 3-3	271	0.238					
Project 8							
injection 4-1	189	0.165	2.3	0.161	0.222	0.070	0.092
injection 4-2	Available						
Total	**4202**	**3.800**	**44.74**	**3.14**	**4.31**	**1.36**	**1.78**

Line A-Condensing turbine results – Medium savings

Extra electricity (MW)	6.835
Reduction in purchased electricity (M$/yr)	2.16
Turbine cost (M$)	10.3

Grassroot – high savings

Project	HX area (m2)	Capital Cost (M$)	LP Steam (t/hr)	Steam Savings (M$/yr)	electrical production drop (MW)	electricity purchased (M$/yr)	Net operating savings (M$/YR)
Project 1							
WW to washing	167	0.101	5.10	0.358	0.500	0.158	0.200
Project 2							
WW to Bleaching	167	0.102					
Project 3							
HW to Bleaching	1023	0.642	12.30	0.863	1.180	0.372	0.491
Below pinch 1	628	0.403					
Below pinch 2	210	0.126					
Above pinch	186	0.112					
Project 4							
HW to Machine	538	0.333					
Below pinch 1	356	0.217					
Below pinch 2	118	0.073					
Above pinch	64	0.043					
Project 5							
HW to Recaust 1	1156	0.755					
Below pinch	308	0.186					
Above pinch	848	0.569					
Project 6							
HW to Recaust 2	1196	0.842					
Below pinch	1126	0.796					
Above pinch	70	0.046					

Project	HX area (m2)	Capital Cost (M$)	LP Steam (t/hr)	Steam Savings (M$/yr)	electrical production drop (MW)	electricity purchased (M$/yr)	Net operating savings (M$/YR)
Project 7							
HW to bleaching form B	910	0.586					
Below Pinch	346	0.210					
Above pinch 1	470	0.292					
Above pinch 2	94	0.084					
Project8							
HW to bleaching from B	114	0.071					
use WBL to digester-36w							
Project 9							
Make up water	579	0.450	9.44	0.663	0.910	0.287	0.376
Below pinch	Available						
Above pinch 1	347	0.310					
Project 10							
Injection 1-3	19	0.017	3.10	0.218	0.180	0.057	0.161
Project 11							
injection 2-1	198	0.119	5.57	0.391	0.330	0.104	0.287
injection 2-2	416	0.379					
injection 2-3	68	0.045					
Project 12							
injection 3-1	198	0.119					
injection 3-2	119	0.074	5.60	0.393	0.330	0.104	0.289
injection 3-3	349	0.313					
Project 13							
injection 4-1	263	0.231	4.20	0.295	0.250	0.079	0.216
injection 4-2	12	0.013					
Project 14							
Boiler Air	1905	1.494	4.80	0.337	0.460	0.145	0.192
Below pinch	1580	1.205	2.06				
Above pinch	324	0.289	2.74				
Total	**9165**	**6.549**	**50.11**	**3.517**	**4.140**	**1.306**	**2.21**

Line A - Condensing turbine results – grassroot high savings

Extra electricity (MW)	7.66
Reduction in purchased electricity (M$/yr)	2.41
Turbine cost (M$)	11.45

3-4 <u>List of modified heat exchanger network and projects for line B – Medium</u>

Project	HX area (m2)	Capital Cost (M$)	LP Steam (t/hr)	Steam Savings (M$/yr)	electrical production drop (MW)	electricity purchased (M$/yr)	Operating savings (M$/yr)
Project 1							
Air Heater	3716	3.520	9.8	0.689	0.585	0.184	0.505
Below pinch	2415	2.071					
Above pinch	1301	1.450					
Project 2							
Make up Water	1347	1.135	20.4	1.429	1.210	0.382	1.047
Below pinch	570	0.362					
Above pinch	776	0.774					
Project 3							
Dilution conveyer NIM	345	0.210	1.2	0.084	0.072	0.023	0.062
CP Indirect cooler	308	0.187					
Project 4							
White water tank NIM	2490	2.074	4.2	0.294	0.250	0.079	0.215
Below pinch	Available						
Above pinch	2332	1.978					
Project 5							
Bleach Heater	1610	1.234	3.3	0.232	0.197	0.062	0.170
Direct Condenser	Available		5.2	0.362	0.307	0.097	0.265

Project	HX area (m2)	Capital Cost (M$)	LP Steam (t/hr)	Steam Savings (M$/yr)	electrical production drop (MW)	electricity purchased (M$/yr)	Operating savings (M$/yr)
Pre projects							
Cold Blow Cooler	1003	0.656					
Below pinch	812	0.541					
Above pinch	190	0.115					
Green liquor cooler	342	0.209					
Below pinch	227	0.137					
Above pinch	115	0.072					
Project 6							
Injection1	418	0.256	5.2	0.361	0.307	0.097	0.265
Injection 1--1	315	0.191					
Injection 1--2	103	0.065					
Project 7							
injection 2	510	0.318	3.3	0.228	0.194	0.061	0.167
Injection 2--1	435	0.269					
Injection 2--2	75	0.049					
Project 8							
injection 3	225	0.141	2.8	0.196	0.166	0.052	0.143
Injection 3-1	112	0.070					
Injection 3-2	112	0.070					
Project 9							
injection 4	235	0.147	3.7	0.260	0.220	0.070	0.190
Injection 4-2	198	0.120					
Injection 4-3	37	0.028					
Total	**12390**	**9.673**	**58.9**	**4.135**	**3.507**	**1.106**	**2.753**

Line B - Condensing turbine results – Medium savings

Extra electricity (MW)	9.00
Reduction in purchased electricity (M$/yr)	2.84
Turbine cost (M$)	13.5